HEAT INTEGRATION STUDY ON BENFIELD AND BUTANE RECOVERY UNIT OF GAS PROCESSING PLANT

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HEAT INTEGRATION STUDY ON CRYOGENIC AND PRODUCT RECOVERY UNIT OF GAS PROCESSING PLANT

By

NGUYEN THI QUYNH NGA

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AS A REQUIREMENT FOR THE

DEGREE OF MASTER OF SCIENCE

IN CHEMICAL ENGINEERING

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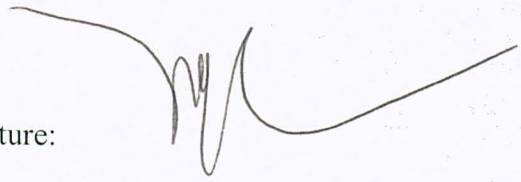
Signature: 

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ABSTRACT

The production and consumption of natural gas is on the rise throughout the world as the result of its wide availability, ease of transportation and use as well as clean burning characteristics. The first and second generation gas processing plants, designed and built in the 1960s and 1970s were bogged down by operational inflexibility and CO_2 freezing due to the rapid change of product values. Even most modern gas processing plants are encountering problems related to operating flexibility, instability of operational conditions due to vapor-liquid equilibrium, CO_2 solid formation and energy efficiency. Ortloff have recently developed new NGL recovery processes on the basis of vapor-split concept. While the Ortloff processes offer relief to the first three problems in gas processing plants, huge energy consumption remains a major concern. A typical gas processing plant consists of two main processing sections, i.e. a cryogenic section and product recovery section. Of the two sections, cryogenic section consumes a huge amount of energy, especially by its cold utility due to the sub-ambient operating condition required. Meanwhile, steam generation for reboiling purposes at the product recovery section also contributes to the huge energy consumption. Ultimately, both result in substantial utility cost to the plant operations; therefore optimizing energy consumption in both areas is essential to improving plant profitability as it results in lower utility cost.

This research looks for improvement opportunities in energy consumption in these two sections based on pinch analysis and maximizing the power generated from turbo-expander. The study explores the utilization of available cold energy extracted from the feed in order to reduce the cold utility requirement in cryogenic area. The study is able to save 6% on refrigeration load for cryogenic area and generate 36% more power from the turbo expander. This could generate predicted annual savings of RM900k for this area. For product recovery area, this study introduces a prefractionator arrangement in place of a conventional depropansier and debutanizer arrangement. Although thermal coupling arrangement requires refrigeration for condenser cooling, nevertheless it generates 54% savings on steam consumption and 37% savings on cooling duty. This is equivalent to RM 2.9mil as predicted annual savings for the plant. Moreover, there are some loose end heat exchangers that could give about RM350k savings per annum for operating cost.

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Chapter 1

INTRODUCTION

1.1. Overview of Natural Gas

The production and consumption of natural gas (NG) is rising throughout the world due to its wide availability, ease of transportation and use as well as clean burning characteristics. Approximately 23 percent of energy consumption in the U.S. comes from NG. Over one-half of homes in the U.S. use NG as their main heating fuel (Energy Information Administration, December 2006). NG is also an essential raw material for many consumer products, e.g. paints, fertilizer, plastics, antifreeze, dyes, photographic film, medicines, and explosives etc.

NG burns cleaner than any other fossil fuels. If NG is of high purity (i.e. methane purity of 95 mole percent or higher), the amount of carbon dioxide (a greenhouse gas) produced is considerably less, due to its lower carbon:hydrogen ratio for methane compared to all other hydrocarbon fuels. Its burning also results in lower emission of sulfur, carbon and nitrogen, as compared to coal or oil. Therefore the use of NG has broadened to cover various industries, especially electricity generation.

The main constituent of NG is methane, a desirable primary fuel. NG also contains small amounts of heavier hydrocarbons e.g. ethane, propane, butane-plus components etc. These heavier hydrocarbons can be recovered profitably in the gas processing plants as liquid products.

Other gases that present in NG include nitrogen, helium, argon, hydrogen and oxygen. Most NG contains some nitrogen and few have 30 mole percent or more. Nitrogen lowers the heat of combustion of the gas. Since NG is normally sold on the basis of energy content with a fixed minimum heating value, its nitrogen contents must be reduced to the lowest possible to meet the specification of the commercial sales gas. Too high nitrogen content may render the NG uncommercial. The removal of nitrogen requires expensive cryogenic processing.

Table 1.1 shows typical compositions of NG, as produced from the reservoirs.

Table 1.1: Constituents of Natural Gas (NG Analysis from Plant X, August 29, 2005)

Component	*Formula*	*Typical composition (mol%)*
Helium	He	-
Argon	Ar	-
Nitrogen	Ni	0.32
Oxygen	O_2	-
Hydrogen Sulphide	H_2S	-
Methane	CH_4	92.90
Carbon Dioxide	CO_2	1.17
Ethane	C_2H_6	3.64
Propane	C_3H_8	1.13
Iso-Butane	iC_4H_{10}	0.22
n-Butane	nC_4H_{10}	0.24
Iso-Pentane	iC_5H_{12}	0.09
n-Pentane	iC_5H_{12}	0.06
Hexane Plus	C_{6+}	0.23
Water	H_2O	-

NG is classified as "sweet" or "sour" based on its hydrogen sulfide and carbon dioxide content. If the content of hydrogen sulfide in NG is low, it is considered "sweet". Conversely, "sour" NG contains high concentration of hydrogen sulfide. In the gas industry, "sweet" or "sour" of NG refers to the presence of both acid gases (CO_2 and H_2S) but usually it refers to hydrogen sulfide alone. Removal of hydrogen sulfide to a very low concentration (4ppmv) or 0.25 g/ 100scf) is required in the field. However carbon dioxide can be tolerated to a much higher level (1-2%) as long as the heating value of NG remains satisfactory (Francis *et. al.*).

Water or brine must also be removed from the NG at the field production facilities, by means of water knockout vessel to prevent water from entering the gas compression/gathering systems. At high pressure and low temperature, NG and free liquid water form solid hydrates which cause blockage in flow lines and downstream equipments.

Transportation of NG can be done using either pipelines (in gaseous form) or tankers (in liquefied/compressed form). Typically, NG is transported from the field to NG customers via gas receiving terminal and the gas-processing plant (GPP). NG transported via pipeline for sales is required to meet contractual sales gas specifications and the typical specifications are shown in Table 1.2:

Table 1.2: Typical Natural Gas Pipeline Specifications (Francis *et. al.*)

Characteristic	*Specifications*
Water content	4-7 lb/ MMscf max
Hydrogen Sulfide content	3.6×10^{-5} lb/ 100scf max
Gross heating value	950 Btu/scf min
Hydrocarbon dew point	15°F at 800 psig max
Mercaptan content	2.9×10^{-5} lb/100scf max
Total sulfur content	$1.4 - 7.1 \times 10^{-4}$ grain/100scf max
Carbon dioxide content	1-3 mole percent max
Oxygen content	0-0.4 mole percent max
Delivery temperature	120°F max (approximately 54°C)
Delivery pressure	700 psig min

These specifications are fixed by negotiation between seller and buyer and vary from case to case. The most important specifications are the first three as both water and hydrogen sulfide must be removed to a very low concentration. Methane itself has relatively low heating value compared to other heavier hydrocarbons. Therefore, enough heavier hydrocarbons are usually present to provide the required heating value.

When chilled to very cold temperatures, i.e. at approximately -162°C, NG changes into liquid form, which is called liquefied natural gas (LNG). Once in this form, it takes only one six-hundredth of the space that it would take in its gaseous states. The ability to store and transport natural gas in LNG tankers has become more important especially where LNG pipeline transport is not economical.

1.2. Natural Gas Processing

Figure 1.1 shows a simplified process flow diagram of a gas processing plant (GPP). This module may be used to condition NG for sales, to extract and recover NGL's or both. The gas processing module may be further divided into sub-modules which are common to many gas processing facilities.

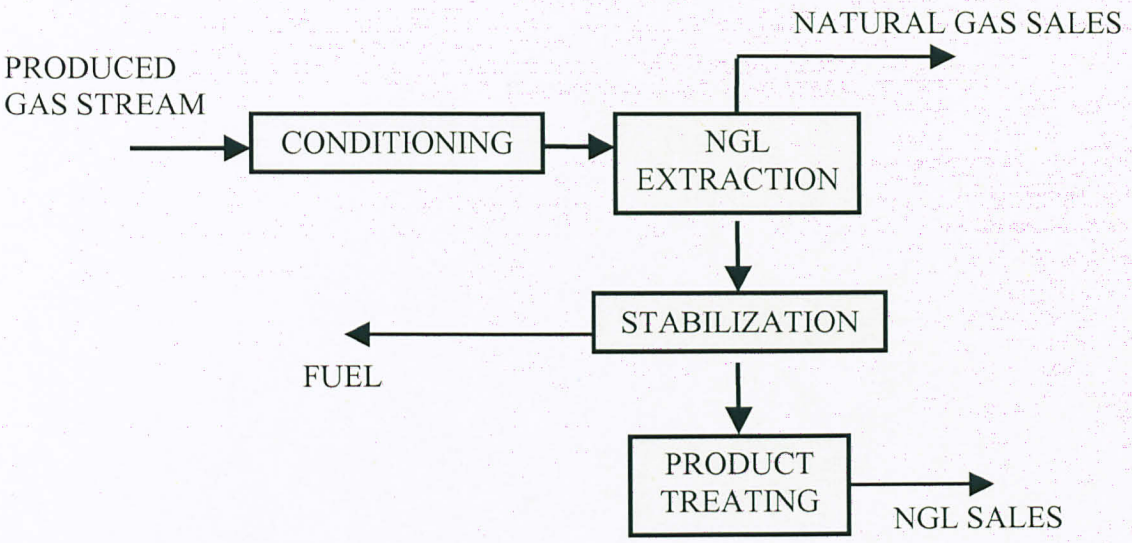


Figure 1.1: Simple flow diagram for an NGL Extraction Plant
(John M. Campbell, Volume 1)

1.2.1. Gas Conditioning Module

The feed gases enter the plant from the surface production facilities that process the gas to various degrees. Depending on the gas processing network and distances from the field, the feed gas may already be dehydrated. In other cases, the feed gas may be wet with possibly two or three phases.

At the downstream of the slug catchers or inlet separators, feed gas is typically filtered and acidic (sour) gases will be treated through amine processing followed by processing in a sulfur-recovery unit. After sweetening, the process gas may be compressed and cooled prior to dehydration and mercury removal.

This module normally consists of the following processes:

- *Gas sweetening process* is to remove H_2S and CO_2 by means of chemical absorption using amine based solvent, physical absorption, solid bed or divert conversion, membranes or extractive distillation. H_2S removal must be essentially complete, while the extent of CO_2 removal depends on the intended use of the gas. If the gas has to be cooled to cryogenic temperature at downstream processes (less than -73°C), CO_2 removal to a few tenths of a percent may be required to prevent formation of solid CO_2 in the cold sections of GPP (John Campbell).

- *Gas dehydration* is to remove H_2O using absorption (tri-ethylene glycol), adsorption (solid desiccant such as silica gel, alumina or molecular sieve), condensation or others such as membrane, $CaCl_2$. Dehydration is often necessary to prevent formation of gas hydrates that may form and plug processing equipment or pipeline at high pressure, even at temperature considerably higher than 0°C. Expander based C2-recovery processes usually require dehydration to less than 5-ppm water (sometimes less than 1 ppm) to avoid water freezing in low temperature process equipment (Francis *et al.*).

- *Cryogenic fractionation* from methane is to remove nitrogen. Operating temperature can be as low as -185°C.

- *Mercury removal* is typically done by passing the gas across a bed of sulfur-impregnated activated charcoal. Ethane-recovery plants usually use aluminum plate-fin heat exchangers because they offer high performance for a relatively low cost. Effluent levels of Hg should be less than 1ppb to avoid damaging aluminum process equipment (John M. Campbell).

The pretreated process gas is now ready to feed the NGL-extraction facility.

1.2.2. Natural Gas Liquids Extraction Module

Three classes of processes for deep recovery of NGL include refrigeration, expander-based cryogenic and lean-oil absorption.

Mechanical refrigeration and lean-oil processes are more competitive when feed gases are rich and available at moderate pressure, typical of those in the Middle East and Nigeria. Basic refrigeration processes are flexible and appropriate for various levels of NGL recovery. Multi-stage C_3 chilling is typically used for C_{3+} recoveries. C_{2+} recovery generally requires lower refrigeration temperatures with mixed refrigerant or cascade cycle processes. These processes, however, are usually not competitive with expander-based processes except for rich or moderate-pressure feeds (20-30 bar).

Expander-based plants tend to dominate C2+ NGL recovery because they can achieve deep recoveries at lower capital and operating expense. The expansion of the pre-cooled feed gas is carried out isentropically across turbine, which recovers useful work from expansion to drive compressor for recompression of residue gas. Gas expansion across turbine generate temperature drop of gas outlet from turbine which is essentially useful for effective NGL separation. These processes dominate NGL recovery, particularly when C_2 or deep C_3 recovery is required.

1.2.3. Stabilization Module

After NGL has been extracted from the feed gas, it must be stabilized to meet sales specifications. These specifications can vary depending on the end markets requirement and transportation method. The stabilization of extracted NGL is accomplished by distillation.

1.2.4. Product Treating Module

After stabilization, NGL must be treated to meet specifications on the contaminants levels (i.e. CO_2, sulfur compounds and water) prior to sale.

1.3. Current issues on Natural Gas Processing

Recovery of NGL from NG is quite common in natural gas processing, and at times can be great economic importance. Huge quantities of solution (associated) gas produced from crude oil used to be flared routinely due to low downstream demand, which results in great economic loss. NGL consists of ethane (C_2), propane (C_3), butane (i-C_4 and n-C_4) and natural gasoline (i-C_{5+}).

1.3.1. Issue 1: Rapid change in product economic values

Prior to 1984, the value of ethane as raw material for petrochemicals was sufficiently high that its recovery was economically feasible. However, the value of sales gas plunged from about $3.50/MMBtu in early 1984 to about $1.50/MMBtu in late 1986 (West-Texas intrastate prices). In 1986, there was very little economic incentive for liquid recovery; therefore, liquid was worth less than its value in sales gas during this period. One factor in the price reversal is the relatively-cheap naphtha, and n-butane, which competes with ethane as the raw material for ethylene production. Ethylene production forms the largest demand for liquid ethane. However, some ethylene plants cannot crack naphtha.

During the Asian crisis of 1997-1998, ethane and its derivatives, ethylene, enjoyed great demand and command high price, whereby the value of ethane in liquefied form had increased from $350/ton in 1996 to $550/ton in 1998, an increase of more than 1.5 times. Therefore, NGL-recovery plants are required to be versatile. The flexibility of operating the process to either recover or reject ethane without sacrificing efficiency or propane recovery is consequently the critical factor in determining the profitability of a GPP. The economic value of liquid propane has always been higher than its gas BTU value; therefore the loss of propane recovery will lead to loss of a plant's profitability.

1.3.2. Issue 2: Instability in operating conditions due to vapor-liquid equilibrium

The second issue of many first generation expander plants is the experience in instability problem associated with vapor liquid equilibrium and carbon dioxide freezing. They are due to the dependence of such plants on using the expander outlet stream to provide column reflux. Since a colder expander inlet temperature produces a colder expander

outlet temperature and generates more liquid in the top column feed, keeping the cold separator temperature as low as possible is necessary to maximize product recovery. Unfortunately in most cases this means operating the separator at high pressure and low temperature, this operating condition falls to a region of the phase envelope for a typical natural gas stream where the vapor-liquid ratio changes quickly. Within the region of phase envelop, small changes in the separator temperature will lead to large change in pressure. When pressure of separator changes significantly, more gas will form and enter the expander, resulting in sudden changes in the expander speed which cause the tower pressure to surge, process gas temperatures to fluctuate. This will go on and on until the whole plant is oscillating. The only way to maintain stable operation is to keep the separator temperature warmer than optimum to avoid this region of instability; however, this will limit the product recoveries than can be achieved. Therefore, the ethane recovery for most of first generation is about 60-70%, which causes a huge loss of ethane to sales gas (Joe Wilkinson *et al., 2003*)

Most GPPs built from the 1970s to the 1990s are based on second generation technology, which introduced external reflux stream and fractionation stages above the expander feed. The second generation technology eliminates dependence on the expander to generate tower reflux and provide better recovery at lower energy consumption than the first generation. However, the second generation processes are often not flexible enough to allow easy adjustment of product recoveries in respond to the rapid changes in product values that are typical in gas processing environment.

Within the liquid recovery section, both operating cost and operating flexibility issues have direct impact on the processing cost. The efficiency of the selected liquid recovery process is an important factor to the processing cost. Moreover, the economic value of ethane in liquefied form fluctuates due to market supply and demand; therefore, the flexibility of operating the process to either recover or reject ethane without sacrificing efficiency or propane recovery is often a critical factor in determining the plant's profitability.

1.3.3. Issue 3: CO_2 solid formation

A third problem for most first generation expander plants is low tolerance for CO_2 in feed gas. CO_2 is a particularly troublesome contaminant often found in natural gas. The content of CO_2 in existing natural gas reservoir varies by geographical locations.

Carbon dioxide falls between methane and ethane in terms of relative volatility. This means that high ethane recovery in a typical NGL recovery plant often suffers high CO_2 concentrations in the NGL product, to the extent that the NGL product must subsequently be treated to remove the CO_2 to meet the purchaser's specification. Although most gas transportation companies and gas consumers will accept CO_2 concentration of several percent in natural gas streams, many NGL recovery processes require removal of the CO_2 to avoid solid formation (freezing) in the cold sections of the processing plant. In most cases, the CO_2 concentration must be kept below 0.5% (often requiring inlet gas treating) to avoid CO_2 buildup in the column. The only alternative is to operate the separator at warmer state so that the column remains warm enough to avoid freezing; however as mentioned earlier, the warmer column temperature shall cause a corresponding loss in product recoveries. Since CO_2 removal equipment will add significantly to both the investment cost and operating cost of the plant's contaminant removal section, there is considerable advantage in using a CO_2 tolerant process in the liquid recovery section of an NGL/LPG recovery facility (Hank *et al.*)

1.3.4. Issue 4: Energy consumption

The fourth issue of a GPP is the energy consumption in NGL extraction area, especially by its cold utility. There, the separation of methane (C1) and ethane-plus components (C2+) takes place. The temperature here can go to as low as -110°C, which requires a huge amount of cold utility, in order to maximize the ethane-plus components recovery as the bottom product in the de-methanizer column. The cold utility is supplied by a refrigeration circuit; the lower the required temperature, the more energy will be consumed by the compression in the refrigeration loop. Therefore, minimizing energy consumption, especially at cold utility, in a GPP becomes a critical factor that contributes to the plant profitability.

1.4. Problem statement

Natural gas processing and consumption is on the rise throughout the world due to its wide availability and use in many industries. Some of the natural gas plants built in 1960s adopted the first generation gas processing plant design, which experience instability problem associated with vapor liquid equilibrium and carbon dioxide freezing. Much work has been done to improve the flexibility of gas processing plant operation, to enable it to better respond to rapid change of product's economic value in order to maximize its profitability. Ortloff have recently developed new NGL recovery processes with high recovery levels, improved efficiency and better CO_2 tolerance on the basis of vapor-split concept (Richard *et al.*, 1998).

Energy efficiency has become the main concern of all plants' operation and gas processing plant is no exception. A typical gas processing plant consists of cryogenic section and recovery section. Cryogenic section consumes a huge amount of energy, especially by its cold utility due to the sub-ambient operating condition. It requires high amount of compression to meet its cooling requirement to a very low temperature. At the recovery section, steam consumption for reboiling purpose contributes significantly to its utility cost. Therefore utility cost becomes the major concern of most natural gas processors in order to improve the plant profitability.

Turbo-expander for gas expansion is widely applied in gas plant in order to recover the expansion energy. The energy generated from the expansion of the gas can be used for sales gas compression, which is able to reduce the gas compression cost. The more energy extracted from the turbo-expander, the more savings can be achieved in terms of compression cost, which contributes to overall plant operating profit.

This research will look for improvement opportunities in energy consumption of these two areas.

1.5. Objectives of the research

The research will focus on the use of pinch technology and simulation tools to produce an energy efficient gas processing plant. The main focus will be on the low temperature separation area (cryogenic area) and product recovery area of gas plant that will:

- maximize the power generated from the turbo-expander in the cryogenic area
- minimize the cold utility consumption in the cryogenic area
- minimize the hot utility consumption in the recovery area

Chapter 2

LITERATURE REVIEW

Natural Gas Liquids (NGL) is a general term applied to liquids recovered from natural gas, i.e. ethane and heavier products. NGL can be extracted by various means. The extraction of NGL is generally preceded by treatment of the gas to remove water, sulfur compounds and other contaminants. Method for the separation of NGL can generally be divided into cryogenic and non-cryogenic systems.

In the early stage, lean oil absorption (non-cryogenic) and mechanical refrigeration were utilized for recovery o f these liquids. The absorption process was later developed into refrigerated absorption process in recent years. The introduction of Joule Thompson valves and turbo expanders in 1960s made significant contributions to the achievement of cryogenic processes. Initially this technology was designed with minimal heat integration and no or little column reflux. These were later developed into schemes that generated column reflux and maximize the heat integration for high NGL recoveries and optimize the plant profitability (Intan Ambari)

Processes such as Turbo-Expansion, Joule Thompson (J-T) Expansion, Cascade Refrigeration and Multi-component Refrigeration (MCR) are commonly referred to as "cryogenic" processes in the gas processing industry. The Turbo-Expander process has been the most attractive due to its deep recoveries at lower operating cost. It is the most practical cryogenic process available to the industry for high ethane recovery, particularly at inlet gas pressure above 500psi (Ewan *et al.*, 1975)

Turbo-expansion recovers the energy of a gas stream after expansion and provides useful cooling. Gas expanded across a valve (J-T) has a certain refrigeration effect, but by expanding the gas across a turbine wheel (turbo-expansion) work is removed from the stream providing a significantly lower final temperature for the same expansion ratio. The lowered temperature both shows a thermodynamic efficiency increase and provides usable energy for mechanical compressor to recompress residue gas from the process.

2.1. The history of gas processing development

First generation NGL/LPG technology is employed in most gas processing plants constructed in the 1960s and 1970s. These designs provide no external expander feed nozzle on the column. This has been referred to as the industry standard single stage or ISS process (Figure 2.1).

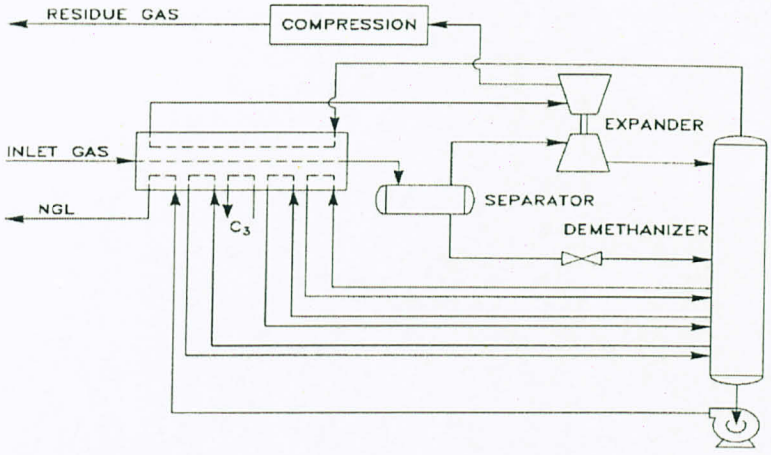


Figure 2.1: First generation of gas processing plant – ISS process (Joe *et al.*, 2003)

Many first generation expander plants purely depend on expander outlet stream as reflux to demethaniser column, which leads to instability in operation associated with vapor liquid equilibrium and carbon dioxide freezing. Since a colder expander inlet temperature produces a colder expander outlet temperature and generates more liquid in the top column feed. Therefore operating separator at temperature as low as possible is essential to maximize product recovery. However in most cases this operating condition will lead the separator to high pressure and low temperature condition, a region of the phase envelope of natural gas where the vapor-liquid ratio changes quickly. In this region, small changes in the separator temperature will lead to large change in pressure, which causes large fluctuation in the amount of vapor entering the expander, resulting large fluctuation in plant operation. Therefore, the only option to avoid this is to operate separator at warmer temperature; however, this will limit the product recoveries that can be achieved.

A second problem for most first generation expander plants is a low tolerance for CO_2 in feed gas. In most cases, the CO_2 concentration must be kept below 0.5% (often requiring

inlet gas treating) to avoid CO_2 buildup in the column that exceeds the solubility limits of the tower liquids, or else CO_2 freezing inside the tower will result. Again the only option is to warm up the separator so that the column remains warm enough to avoid freezing; however the warmer column temperature lead to a corresponding loss in product recoveries (Joe *et al., 2003).*

The second generation NGL/LPG technology was developed to address both the limitations of the first generation processes. The second generation introduced the external reflux stream and fractionation stages above the expander feed to eliminate the operation instability associated with vapor liquid equilibrium and provide better recovery at lower energy consumption.

However, both first and second generation expander plants is not flexible enough to allow adjustment of either recover or reject ethane operation because rejecting ethane in both generations will cause a penalty in propane recovery. Whereas the price of liquid ethane is based almost entirely on its value as a petrochemical feedstock, the value of propane as both fuel and feedstock makes its price in liquefied form more stable. This means that propane is nearly always more valuable as a liquid than as BTUs in the plant residue gas, so losing propane recovery to residue gas in order to reject ethane can cause loss for the plant profitability when liquid ethane market is depressed.

Therefore, the third generation NGL/LPG technology was developed to add flexibility to expander plants for high propane recovery regardless of the ethane recovery by providing two external reflux streams for the column. When operating at ethane recovery mode, both reflux streams are placed above the expander feed; while one reflux stream placed above and one below the expander feed for propane recovery operation mode. In Figure 2.2, the design is for ethane rejection operating mode where the reflux stream by expander outlet is placed between two external reflux streams for the de-ethaniser column.

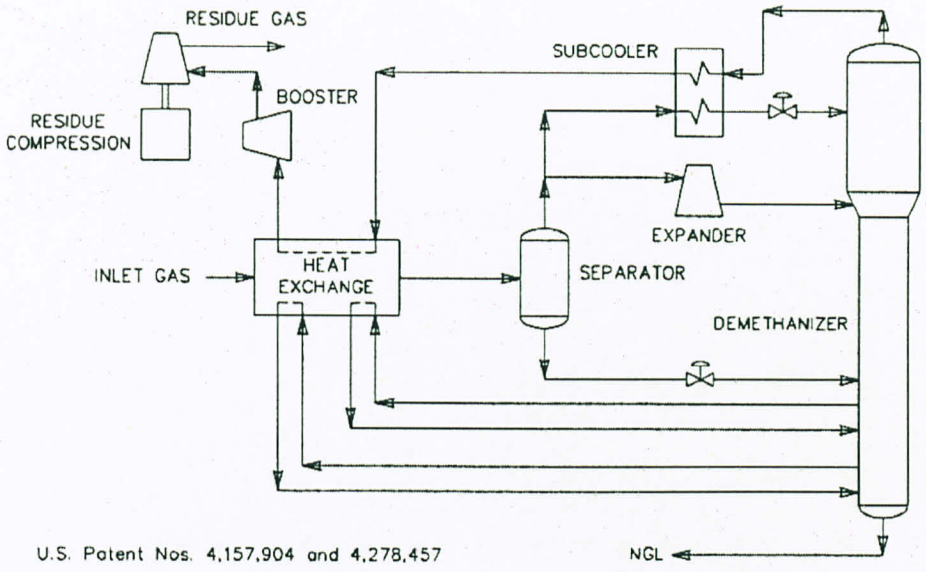


Figure 2.3: Gas-Subcooled Process (GSP) (Joe Lynch, 1996)

When CO_2 is present in the feed gas, the subcooled liquid will absorb most of CO_2 together with C_{2+} components to prevent CO_2 concentrating in the upper, colder sections of the tower, which give higher ethane recovery levels without CO_2 freezing in the column. However, this process is not efficient to operate at ethane rejection mode as propane recovery efficiency suffers significantly due mainly to the higher concentration of propane present in the top feed.

The OverHead Recycle Process (OHR) has been used instead of GSP for LPG recovery plants (Figure 2.4). In this process a vapor stream is withdrawn from an intermediate point in the composite distillation tower (de-ethansier and absorber towers) that is then condensed by overhead stream from absorber column. The condensed liquid is introduced at top stage as reflux for the upper portion of the composite tower. In the absorber, the condensed liquid contacts and rectifies the vapor outlet from the expander, absorbing the C_{3+} components for recovery in the bottom product from the second column.

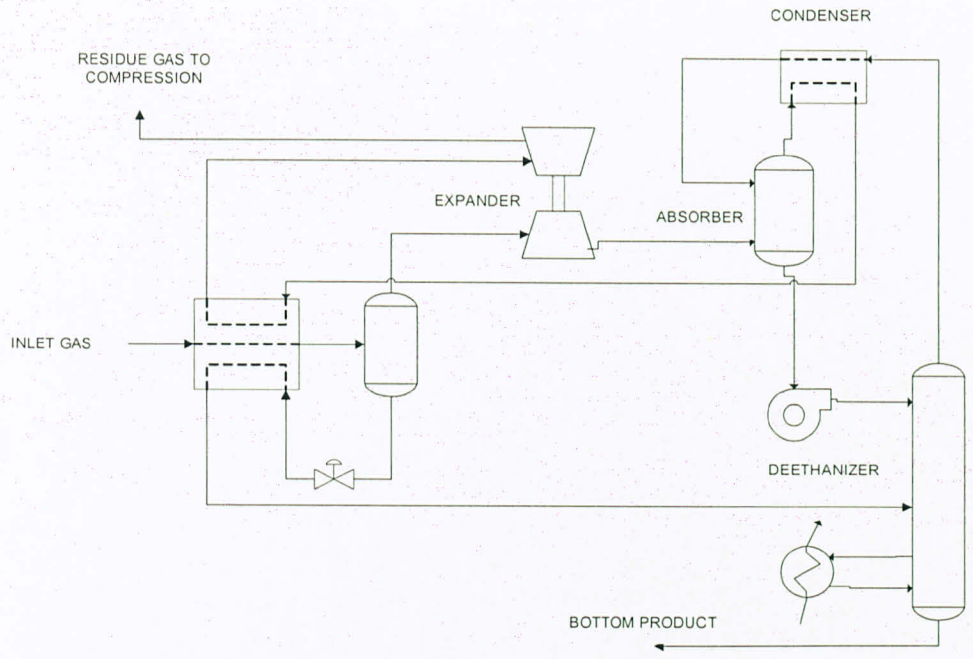


Figure 2.4: Overhead Recycle (OHR) Process (Richard *et al.*, 1998)

This process provides more efficient recovery of propane and heavier hydrocarbons than GSP design but it is not suitable for high ethane recovery.

The split vapor concept has been explored to extend its capabilities to overcome the equilibrium limitations of current processes. Many developments have been done to improve the plant operation and plant profitability. The first improvement is the Cold Residue Reflux process (CRR), which indirectly utilizes the available refrigeration source in the sub-cooled split-vapor stream (Figure 2.5). This process adopts the same split-vapor concept as GSP process while creating a pure methane as reflux stream above the split vapor reflux stream. The pure methane stream needs to be liquefied before being fed to the demethaniser column as reflux. It will be condensed by heat exchanging with the split vapor stream. However, the split vapor stream is not cold enough to liquefy pure methane stream at the demethaniser operating pressure. Therefore, a small compressor is needed to boost a portion of the tower overhead to higher pressure so that it can be condensed by the split vapor stream.

In the demethaniser column, most of ethane plus components will be absorbed by the split vapor stream then the residue ethane will be further absorbed by the condensed pure methane at the top section of the tower.

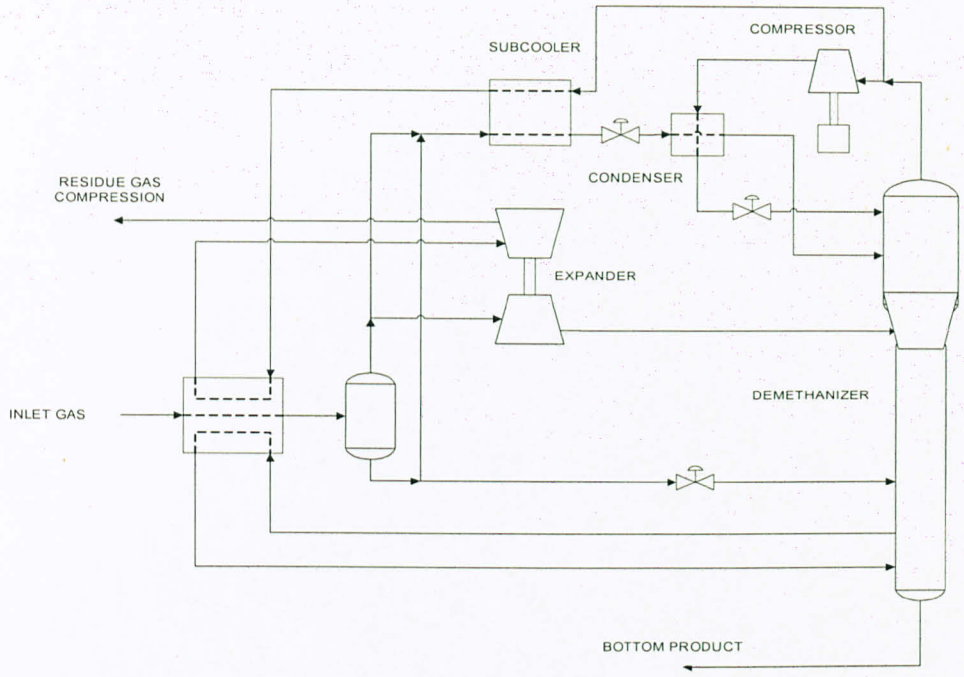


Figure 2.5: Cold Residue Reflux (CRR) process (Richard *et al.*, 1998)

The methane reflux stream could rectify the tower vapors so that it could strip most of the ethane and heavier component from the tower overhead product. Therefore, ethane recovery can reach up to 99% by using this process with the same compression horse power as the GSP design. This design retains the advantage of the GSP design in terms of high ethane recovery without CO_2 freezing in the tower; moreover, it also has ability to operate at nearly complete ethane rejection while maintaining propane recovery in excess of 99% (Richard *et al.*,1998)

Although the CRR process gives the advantage of high recovery efficiency and better CO_2 tolerance, it requires high capital investment due to investment on separate compressor (for compression of pure methane as reflux stream); therefore, high compression horsepower for a given recovery level. The Recycle Split-Vapor (RSV) process is developed to overcome the shortcomings of the CRR process (Figure 2.6).

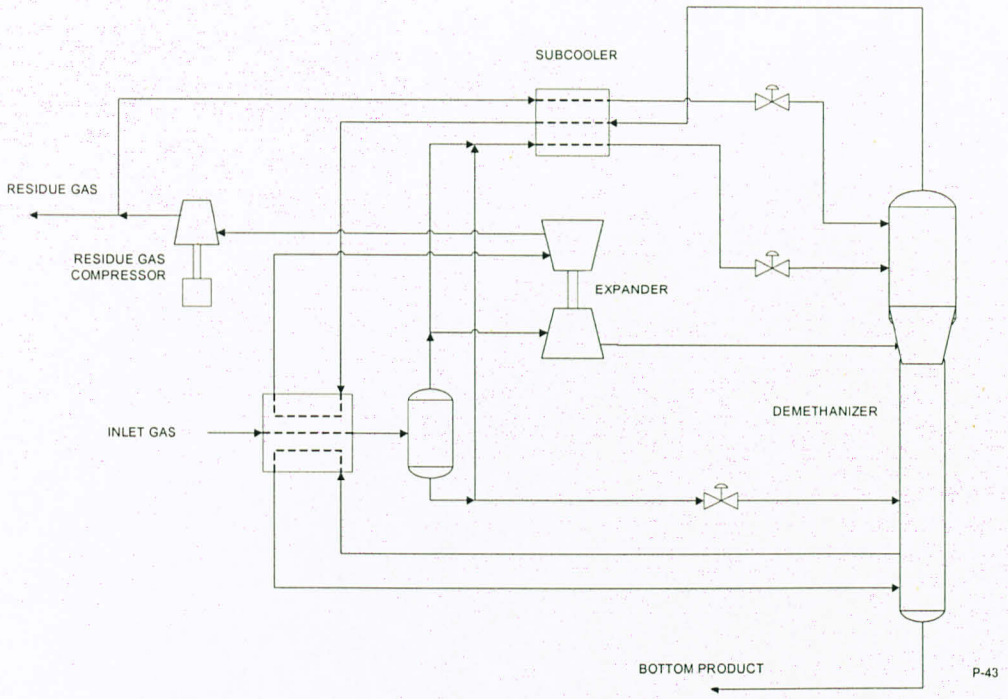


Figure 2.6: The Recycle Split – Vapor (RSV) Process (Richard *et al.*, 1998)

In RSV process, instead of using a separate compressor, a small portion of the recompressed residue gas instead of tower overhead stream is introduced as reflux stream for the tower. The recompressed residue gas is condensed, sub-cooled and then flashed down to tower pressure and supplied as the top feed to the de-methanizer. This stream pressure is high enough to be liquefied by the tower overhead gas stream and the split vapor feed.

This process requires lower compression horsepower for a given recovery level because the split-vapor stream is able to absorb most of the ethane contained in the expander outlet vapor so that the much smaller reflux flow is needed to rectify the residue ethane from the vapors in the upper section of the tower. (Richard *et al.*, 1998)

Both CRR and RSV process are suited for both ethane recovery and ethane rejection operation and can easily switch between the two operating modes to adapt to changes in market price of ethane.

Instead of a separate exchanger required in CRR process or a separate exchanger passage in RSV process, the recompressed residue gas can be mixed with the split-vapor feed

before being condensed and sub-cooled, which could reduce the capital investment. This concept has been applied in the Recycle Split-Vapor with Enrichment (RSVE) process shown in Figure 2.7.

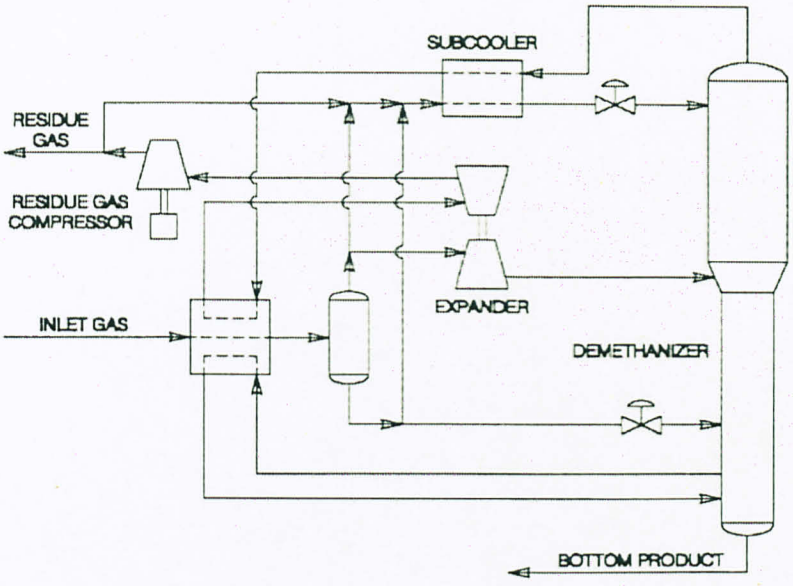


Figure 2.7: The Recycle Split-Vapor with Enrichment (RSVE) process (Richard *et al. 1998)*

Due to mixing of residue gas and split vapor feed, ethane content in the mixing stream is richer compared to the recompressed residue gas. Although this mixed stream as reflux to the top section of the column limits the ethane recovery to a slightly lower level compared to RSV process due to equilibrium effect, the lower capital investment and simplicity in the operation of RSVE as compared with RSV may offset the small loss in ethane recovery in some cases. Moreover, RSVE has better CO_2 tolerance compared to CRR and RSV designs operating at the same ethane recovery level (Richard *et al*, 1998). Enriching the recycle stream with heavier hydrocarbons in the split vapor feed will cause the bubble point temperature raise of the liquid in the upper section of the de-methanizer, which leads to the tower operating at the conditions away from the condition of where solid CO_2 begins to form. Therefore, RSVE design can tolerate significantly higher CO_2 concentrations in the feed gas for a given level of ethane recovery, which makes it the most CO_2-tolerant process yet (Richard *et al*,1998).

2.2.2. Propane Recovery Processes

While the value of propane as both fuel and petrochemical feedstock makes its price as a liquid more valuable than as BTUs in the plant residue gas, the price of liquid ethane is based almost entirely on its value as a petrochemical feedstock. Therefore, when the value of liquid ethane is low, efficient ethane rejection without sacrificing propane recovery is the key to plant profitability.

In order to utilize the available refrigeration source of split-vapor stream, the flashed split-vapor stream is used to cool the tower overhead and generate reflux for the column as shown in the Split-Flow Reflux (SFR) process (Figure 2.8).

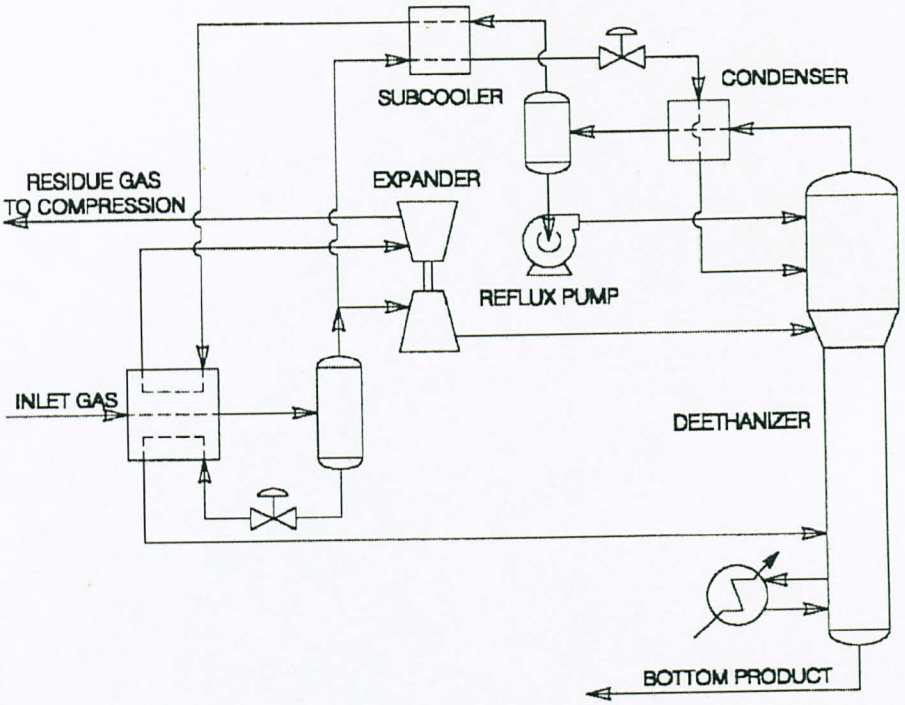


Figure 2.8: The Split-Flow Reflux (SFR) process (Richard *et al.*, 1998)

Due to ethane rejection process characteristics, the tower overhead contains most of ethane and methane in the feed gas; therefore it condenses at higher temperature compared to CRR process. Therefore, the split vapor stream is able to liquefy the tower overhead stream. The liquid condensed from the overhead is then separated via a phase separator and returned as reflux to top of the tower at the higher stage than the split-vapor stream to provide final rectification of the tower vapors for better propane recovery. The

split vapor stream will absorb most of propane at the lower stage. Therefore, a small amount of reflux from the condensed overhead liquid is required to give the final rectification at a given propane recovery.

In order to make a better use of the refrigeration available in feed streams, the Improved Overhead Recycle (IOR) process has been introduced as shown in Figure 2.9.

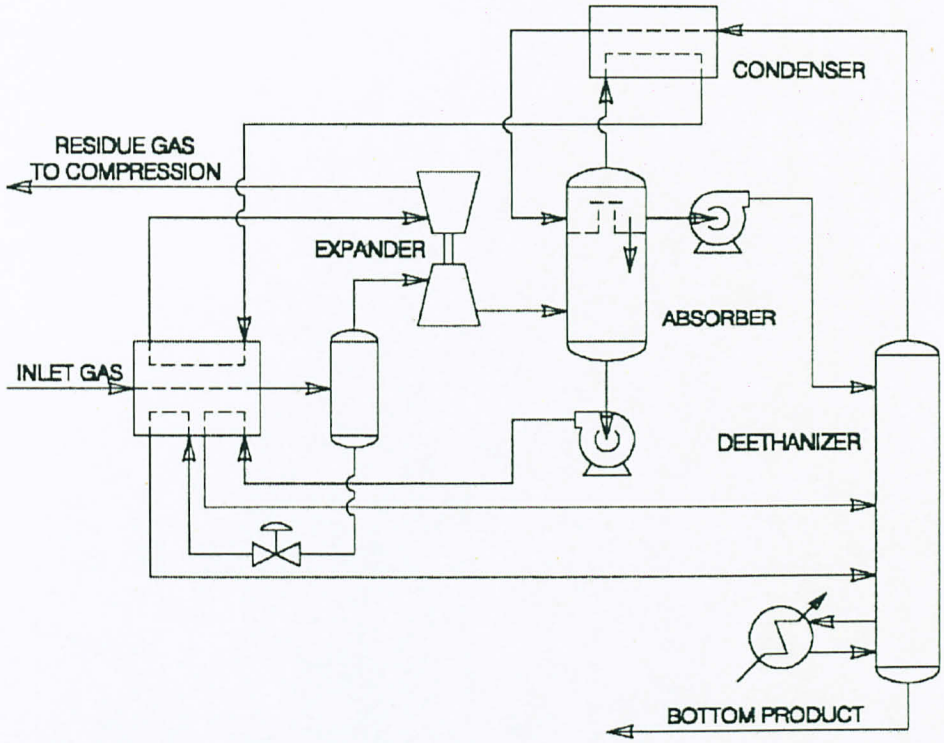


Figure 2.9: The Improved Overhead Recycle (IOR) process (Richard *et al.*, 1998)

Instead of supplying the cold absorber bottoms liquid directly to the de-ethaniser as in OHR process, IOR process utilizes this cold absorber bottoms liquid stream to supply part of the feed gas cooling. This not only reduces the cooling load for the front end of the plant, but also reduces the de-ethaniser reboiler duty due to being partially vaporized before fed to the column.

Both OHR and IOR process have traditionally been employed as two-column systems, the two columns in either process can be visualized conceptually as a single composite column with an intermediate vapor side-draw. This composite column concept led to the development of the Single Column Recycle process (SCORE) shown in Figure 2.10.

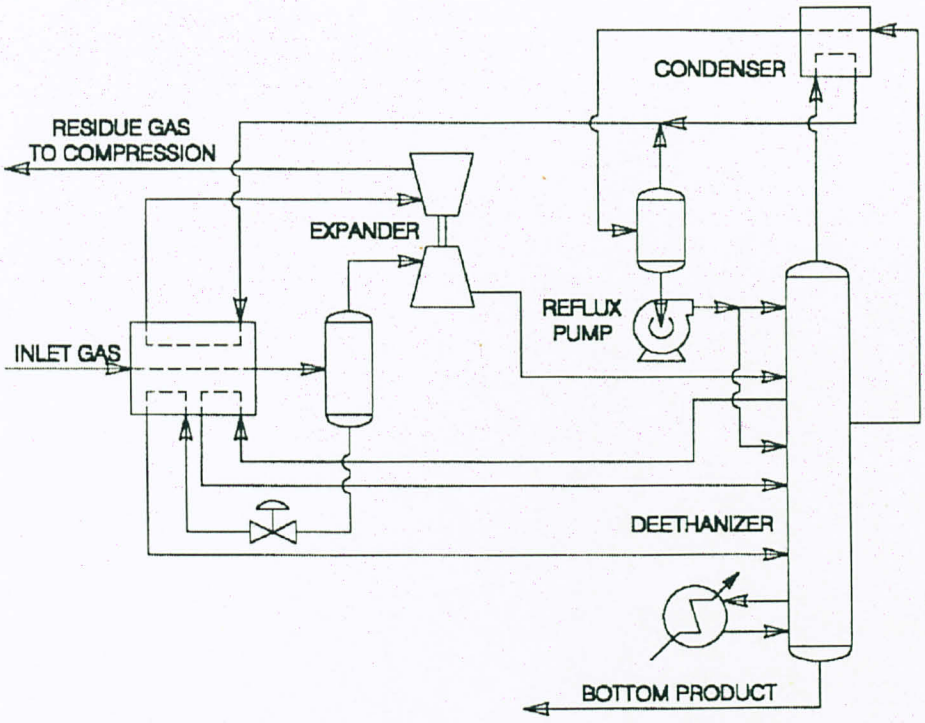


Figure 2.10: The Single Column Recycle process (SCORE) (Richard *et al.*, 1998)

This process design has a considerable advantage in terms of the investment cost for the plant due to the investment for a single, larger column and small reflux drum is generally less expensive than that for the two columns used in both OHR and IOR process, and one less set of cryogenic pumps is required. The single column design is also more easily adapted to ethane recovery operation. In terms of efficiency improvement, the SFR and SCORE designs offer higher recovery for a given amount of compression, less compression for a given recovery level or a combination of both (Richard *et al.*, 1998).

2.3. Comparison of benefits of new processes

Several comparison cases have been done to illustrate the advantages of the split-vapor process designs over the ISS plant. One of them is the work of John D. Wilkinson and Hank M. Hudson published in 1992. Process simulation were performed for a typical inlet gas stream at a flow rate of 100MMSCFD, assuming the plant operates on a nominal 1050 psig pipeline. The comparison was based on percentage of recoveries of ethane, propane and butane liquid products and residue gas compression horsepower for both ethane recovery and rejection. (Table 2.1)

Table 2.1: New plant process comparison (John Wilkinson and Hank Hudson, 1992)

	ISS	GSP	SFR	CRR
Ethane recovery operation mode				
C_2 recovery %	79.6	88.0	88.0	98.0
C_3 recovery %	96.9	97.8	97.8	100.0
C_4 recovery %	99.4	99.4	99.4	100.0
Residue HP	5,774	4,597	4,597	4,732
Incremental kW	0	0	0	158
Ethane Rejection operation mode				
C_2 recovery %	0.8	0.9	0.9	0.9
C_3 recovery %	84.0	93.7	99.9	100.0
C_4 recovery %	96.7	98.6	100.0	100.0
Residue HP	5,783	4,590	4,598	4,731
Incremental kW	0	0	4	158

From the results shown, the three split-vapor feed processes offer notable improvement in ethane recovery and also consume 20% less residue gas compression horsepower for ethane recovery operation mode. For ethane rejection mode, although ISS plant could reject almost all of ethane, it suffers propane loss due to lack of reflux for the tower. Besides that, the three split-vapor feed designs offer substantially higher propane recovery as well as efficient ethane rejection. Moreover, the new revolution plants consume 18-20% less compression horsepower compared to ISS plant. Lower compression horsepower of the new designs also markedly improves the economics for gas plant operation (John *et al*, 1992)

The comparison does not stop at the new plant evaluation. The comparison made on plant retrofit shows the strength of split-vapor feed designs not only on new grass roots plants but also on revamping older, existing gas plants. Figure 2.11 shows the retrofit of ISS plant to GSP design.

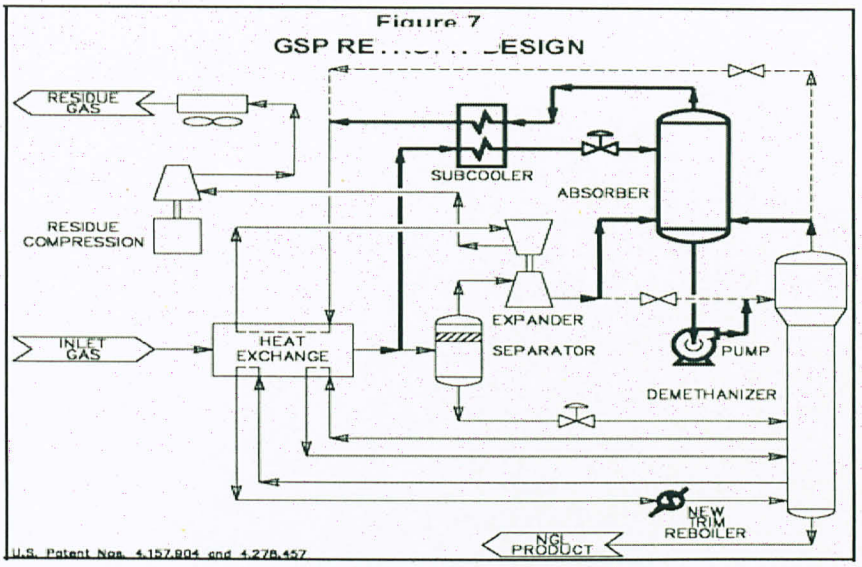


Figure 2.11: GSP Retrofit plant (John Wilkinson and Hank Hudson, 1992)

The thicker lines in Figure 2.11 indicate the modification of existing ISS plant to become GSP plant. For GSP retrofit, a new absorber column is often used to serve as rectification section of demethaniser so that no modification to the existing column is required. Cryogenic pump is required to transfer liquids from the bottom of the new absorber back to the top of the existing demethaniser using the existing expander feed line. However, the pump can be eliminated in some case if the existing demethaniser overhead is allowed to bypass the rectification section, which will in return affect the recovery level of ethane and propane.

Similarly for SFR and CRR retrofit plants from ISS plants, it is shown in Figure 2.12.

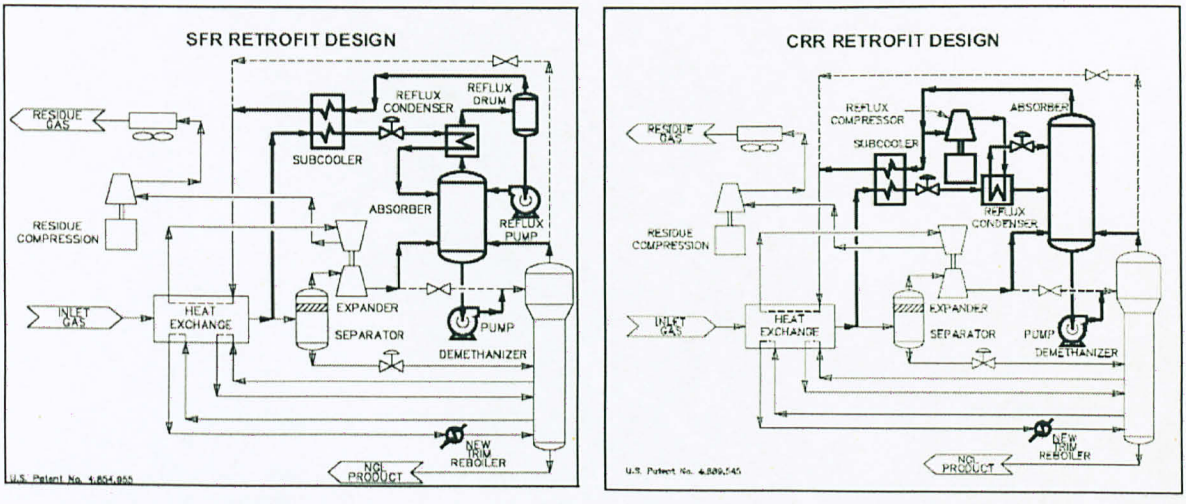


Figure 2.12: SFR and CRR Retrofit Design (John Wilkinson and Hank Hudson, 1992)

The comparison of retrofit plants with existing ISS plant is shown in Table 2.2.

Table 2.2: Retrofit Plant Process Comparison (John Wilkinson and Hank Hudson, 1992)

	ISS	GSP	SFR	CRR
Ethane recovery operation mode				
C_2 recovery %	75.0	88.2	88.2	97.0
C_3 recovery %	96.6	97.9	97.9	100.0
C_4 recovery %	99.4	99.5	99.5	100.0
Residue HP	6,000	4,811	4,811	4,976
Incremental kW	0	0	0	145
Ethane Rejection operation mode				
C_2 recovery %	40.0	0.9	0.9	0.9
C_3 recovery %	85.2	94.0	99.9	99.9
C_4 recovery %	97.1	98.7	100.0	100.0
Residue HP	5,212	5,102	5,118	5,096
Incremental kW	0	3	4	132

From the results shown for ethane recovery mode, the ISS plant gives 75% of ethane recovery with 6000 HP, while the retrofit designs provide increased ethane recovery with 20% less compression horsepower consumption. For ethane rejection mode, the ISS plant limits the amount of ethane rejected in order to maintain the propane recovery at reasonable level. However, the retrofit design can fully reject ethane with a substantial efficiency in propane recovery compared to ISS design. (John *et al*, 1992)

2.4. LPG Recovery from Natural Gas

Natural gas typically contains up to 15 vol % of hydrocarbons heavier than methane which are separated to provide pipeline quality methane and recovered liquid hydrocarbons. These valuable natural gas liquids comprise primarily ethane, propane, butane, and minor amounts of other light hydrocarbons which are recovered from natural gas following compression and initial removal of non-hydrocarbon acid gases, water, and other impurities. The market price of LPG (Liquefied Petroleum Gases), and the constituent products (i.e. Ethane, Propane, Butane and Pentane plus), the gas composition, and the Plant Capital and Operating Costs determine which process to choose and the percent recovery of each component required.

Depending on the local market requirements, the LPG recovered can be converted into a variety of products, including:

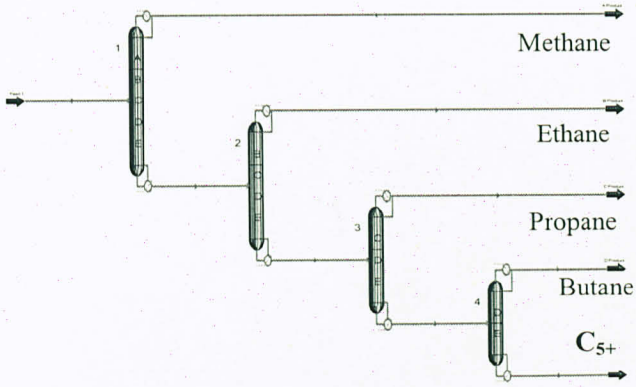
- C3+ Product
- Stabilized Condensate
- Ethane
- Commercial Propane or HD-5 Propane
- Commercial Propane/Butane Mix
- Commercial Butane

A typical NGL product contains approximately 40 vol % C_2, 40 vol % C_3, and 20 vol % C_4 and heavier hydrocarbons; actual NGL composition varies with the natural gas source. Depending on the products required, some or all of the following systems will be required:

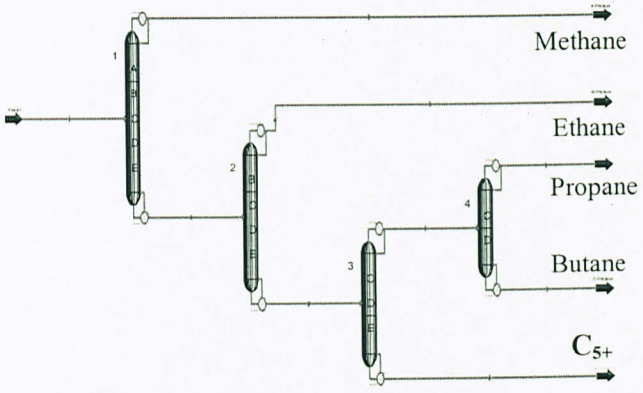
- De-methanizer
- De-ethanizer
- De-propanizer
- De-butanizer

Most of the industrial practice has been resorted to direct or indirect sequence. Direct sequence is the process first to recover ethane (de-ethaniser column), followed by propane (de-propaniser column), butane (de-butaniser column) and pentane-plus components consequently; whereby indirect sequence is the process to first separate ethane (de-ethaniser column), followed by separation of pentane-plus components from

propane and butane (LPG column) and lastly separate propane from butane (de-propaniser column) as shown in Figure 2.13.



a. Direct sequence



b. Indirect sequence

Figure 2.13: Direct and Indirect sequence for LPG recovery

Theoretically, direct separation sequence is normally more favorable in most of gas processing plants as it normally requires less energy consumption.

2.5. Summary of Literature Review

From the literature review, it can be realized that most of the previous work is focused in the improvement of plant flexibility to adapt to the rapid change of the market price of ethane as well as to enhance the CO_2 tolerance for the plant operation due to higher and higher CO_2 content in the natural gas source. Many improvements have been done in order to improve product recovery efficiency to adapt to change in gas plant margin.

Energy consumption is also a critical concern in gas processing field due to high requirement of energy consumption in gas processing. This research will be focused on improvement in energy efficiency for the low temperature separation unit and product recovery unit, which require huge energy consumption in the gas processing plant.

Chapter 3

THEORY

3.1. Pinch Analysis

In the late 1970s Pinch Technology emerged as a tool for the design of heat exchanger networks. Its key contribution was to give the engineer simple concepts which were used interactively. In mid 1980s, Gundersen and Naess published a seminal review of heat exchanger network design methodology, which contributed Pinch Technology become more widely use in heat exchanger network design. Moreover, the methodology has been extended to address diverse range of systems including distillation, heat pumps, co-generating turbines, furnaces, etc. and to address non-energy objectives such as capital costs, operability and emissions. It is no longer restricted to heat exchanger network, but rather used as a general methodology for conceptual process design upstream prior to flowsheeting and simulation. A general approach has emerged for integrated design of energy and process system. Morgan explained that today's experienced practitioner will use Pinch Analysis to scope and screen alternative options early during the conceptual design of integrated systems (Morgan)

The best known concept in pinch analysis is the "Composite Curves". The Composite Curves are constructed from "stream data" representing a process heat and material balance. It consists of composite hot stream and composite cold stream. The composite hot stream is a single stream that is equivalent to the individual hot streams in terms of temperature and enthalpy; so is the composite cold stream. In the composite curve, the hot composite curve overlaps the cold composite curve, which represents the heat recovered from the hot composite curve to the cold composite curve (Figure 3.1). The more the overlap area, the more heat can be transferred from the hot streams to cold streams; hence, more heat can be recovered. Maximizing the energy recovery will minimize the external requirements for heating and cooling duties and minimize the energy consumption. Heat recovery is only possible in the overlap area; when the cold composite curve extends beyond the start of the hot composite curve, the cold composite curve must be supplied with external hot utility such as steam. Similarly, when the hot composite curve extends beyond the start of the cold composite curve, the hot composite curve must be supplied with external cold utility such as cooling water or refrigeration.

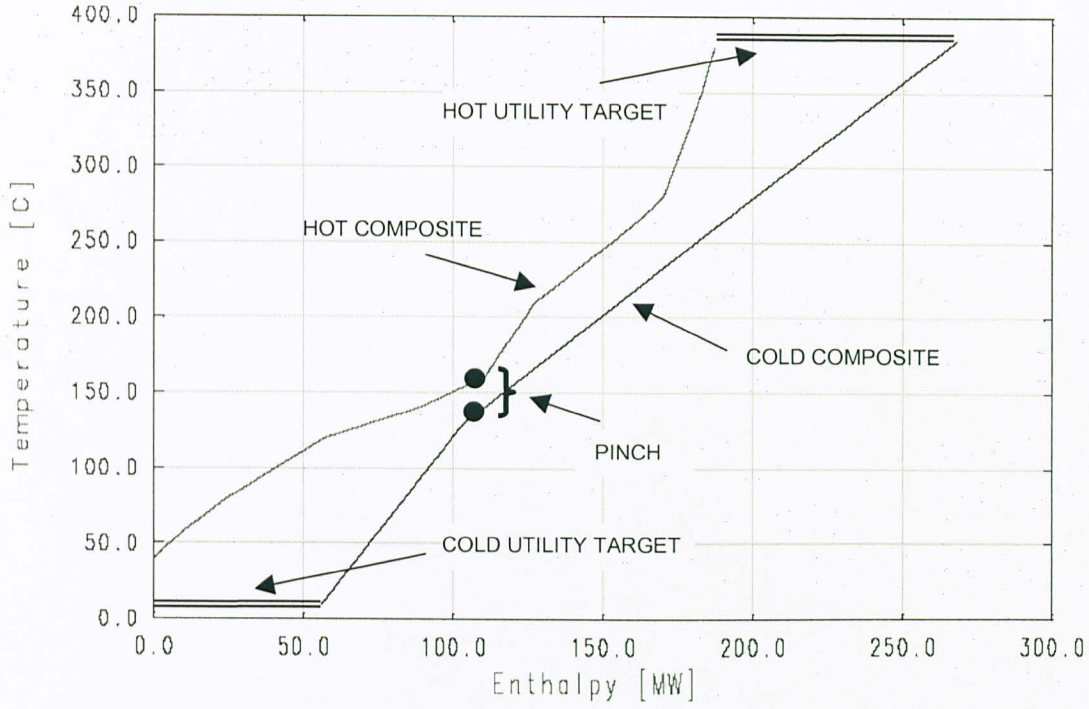


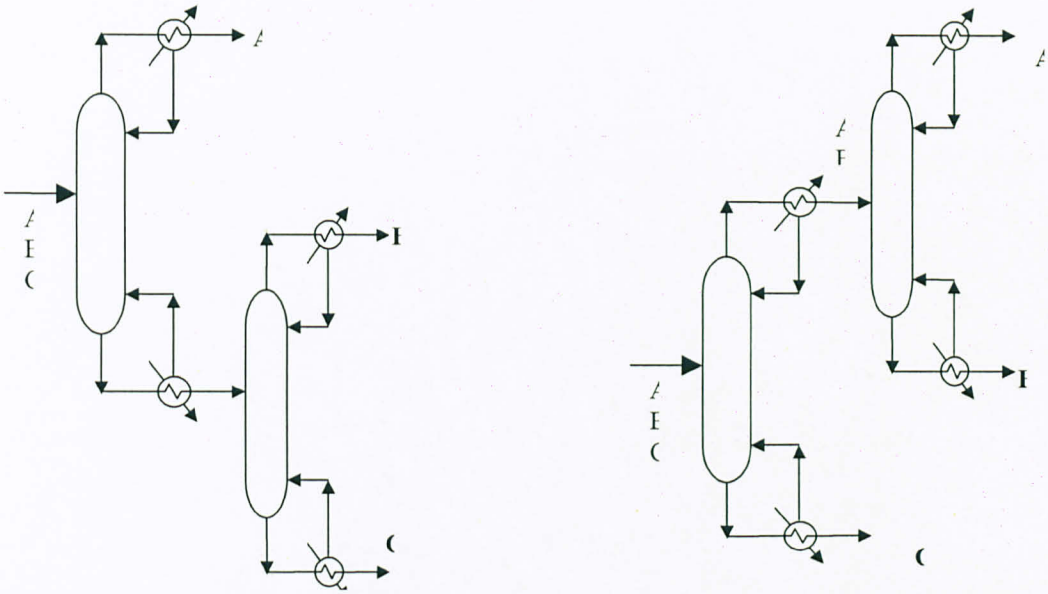
Figure 3.1: Composite Curve with hot and cold utility targets

The composite curve allows the designer to predict optimized-hot and cold utility targets ahead of design, to understand driving forces for heat transfer and to locate the heat recovery pinch. It is also used to visualize heat transfer (sources, sinks, driving forces) in the overall context.

Further details of pinch analysis such as the pinch concept, grid diagram, grand composite curve etc. can be referred to the Pinch Analysis – A-state-of-art Overview, a well-known work of B. Linnhoff, 1993.

3.2. Prefractionator arrangement

When separating a three-product mixture using simple columns, there are only two possible sequences, Figure 3.2.



(a) Direct sequence

(b) Indirect sequence

Figure 3.2: The direct and indirect sequences of simple distillation columns for three-component separation

Moreover, the possibilities shown in Figure 3.3 can be considered.

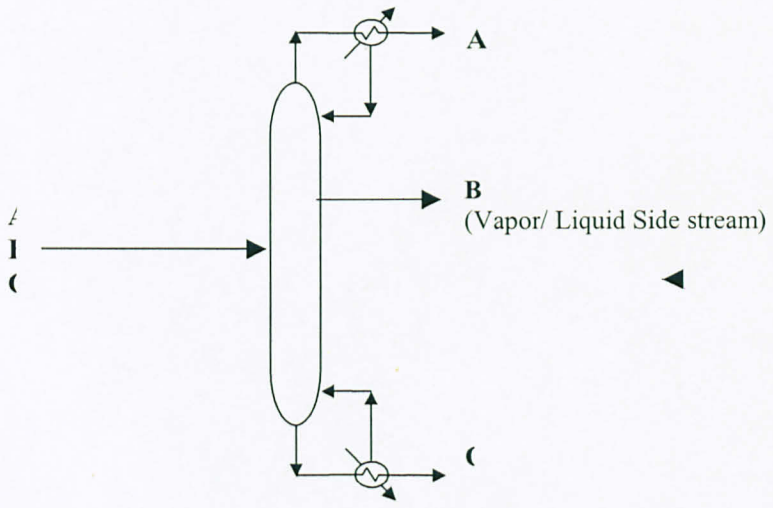


Figure 3.3: Distillation column with three products (Robin Smith)

Here three products are taken from one column. The designs are in fact both feasible and cost-effective when compared to simple arrangements on a stand-alone basis (i.e. reboilers and condensers operating on utilities) for certain ranges of conditions (Robin Smith)

Consider a three-product separation in which the lightest and heaviest components are chosen to be the key separation in the first column. Two further columns are required to product pure products, Figure 3.4a. This arrangement is known as distributed distillation or sloppy distillation.

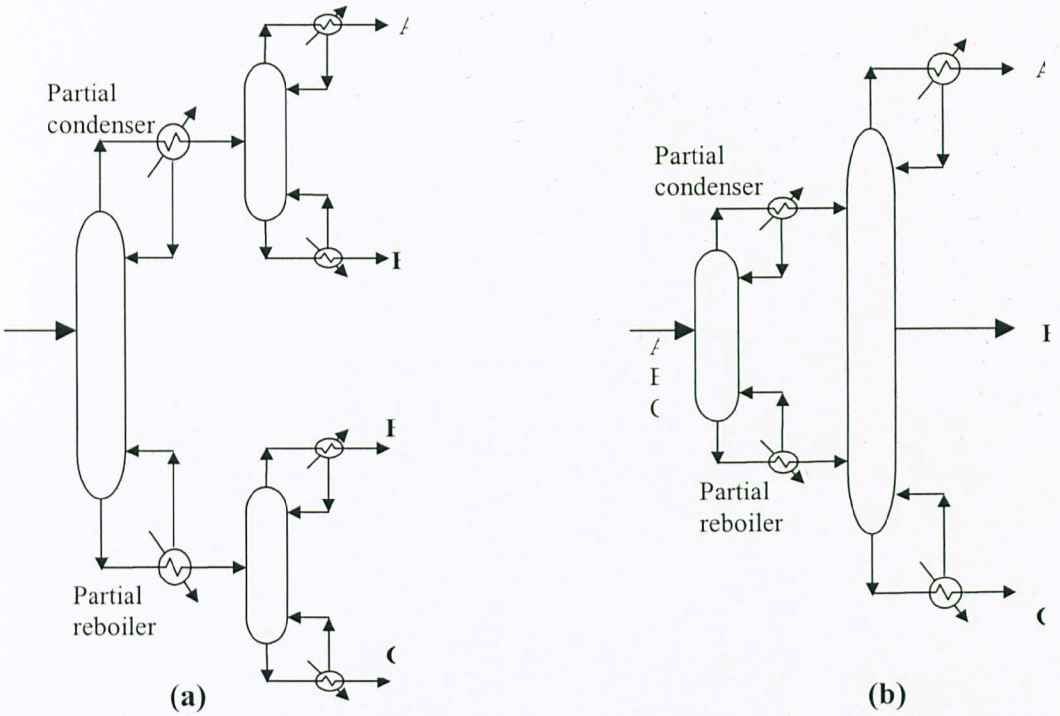


Figure 3.4: Prefractionator arrangement (Robin Smith)

If the second and third columns in Figure 3.4a are operating at the same pressure, then the second and third columns could simply be connected and the middle product taken as a side-stream as shown in Figure 3.4b. This arrangement is known as a prefractionator arrangement. Comparing the prefractionator arrangement with the conventional arrangement in Figure 3.2, the prefractionator arrangement typically requires 20 to 30% less energy than conventional arrangement for the same separation duty due to the more thermodynamic efficiency in the prefractionator arrangement (Robin Smith).

In the direct sequence shown in Figure 3.5, the composition of component B in the first column increases below the feed as the more volatile component A decreases. However, moving further down the column, the composition of component B decreases again as the composition of the less-volatile component C increases. Thus the composition of B reaches a peak only to be remixed (Robin Smith).

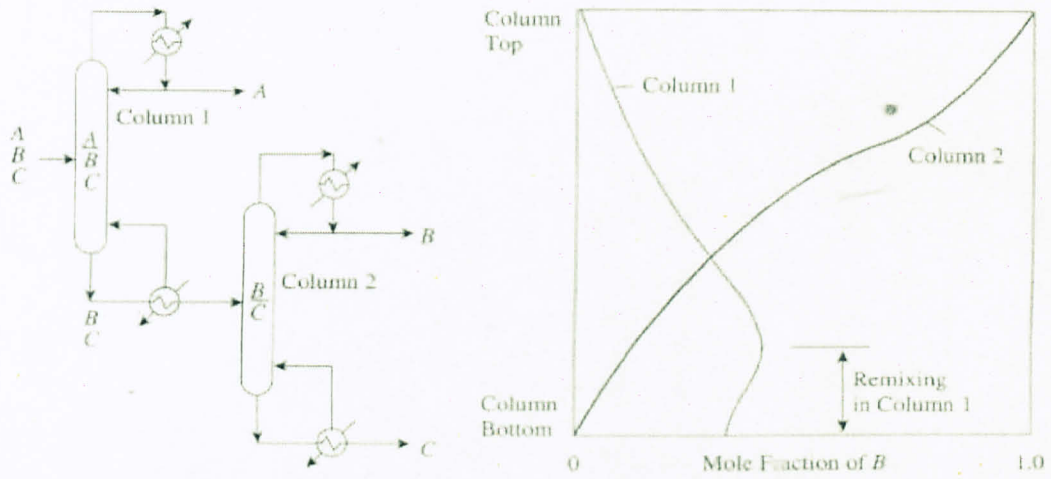


Figure 3.5: Composition Profiles for the middle product in the columns of the direct sequence showing remixing effects (Robin Smith)

This remixing is a source of inefficiency in the separation. However, this remixing can be avoided in the prefractionator arrangement shown in Figure 3.6.

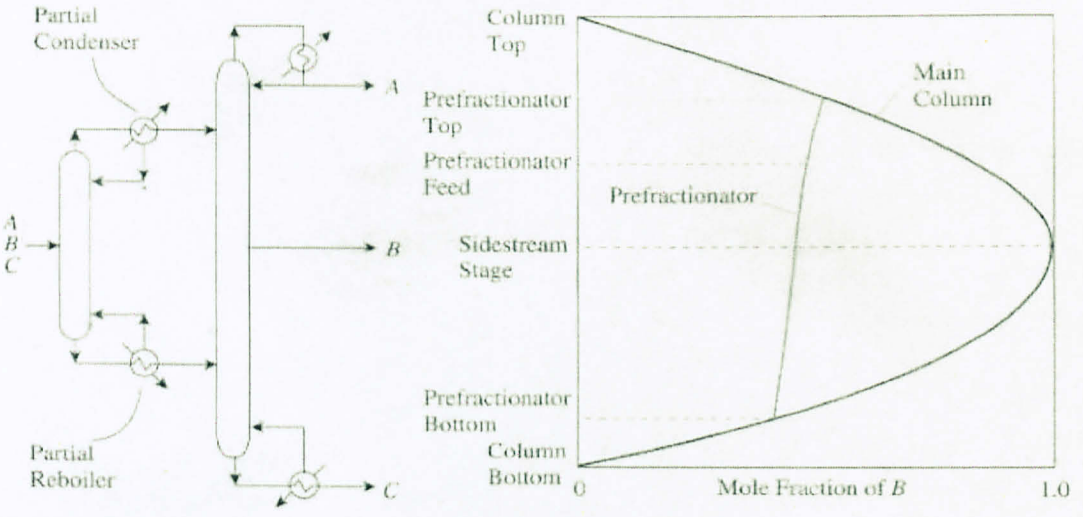


Figure 3.6: Composition profiles for the middle product in the prefractionator arrangement showing no remixing effects (Robin Smith)

In the prefractionator, the remixing effects are avoided because each sections of the column only remove one component from the product mixture of that section. The upper section of the prefractionator separates AB from C; whilst the lower section separates BC from A. This is also applied for all four sections of the main column. In this way, the remixing effects are avoided (Robin Smith).

3.3. Turboexpander

Turboexpander is a centrifugal or axial flow turbine through which a high pressure gas is expanded to produce work that is typically used to drive a compressor. Because work is extracted from the expanding high pressure gas, the expansion is isentropic and the low pressure exhaust gas from the turbine is at a very low temperature, often as low as 200 K (-73°C) or less (Ralph James *et al.*).

Turbo-expanders are used in all gas plants. It is a remarkable turbine-like device used to cool plant inlet gas by reducing its pressure. The natural gas is cooled to extremely low temperatures through a heat exchanger network after which the cold liquid and vapor are separated in a low temperature separator. The liquid stream is then flashed across a Joule-Thompson (JT) valve for pressure reduction and additional cooling. The vapor stream of the separator is fed to the expander side of turbo-expander where temperature is further reduced and the work produced is utilized for recompression of sale gas.

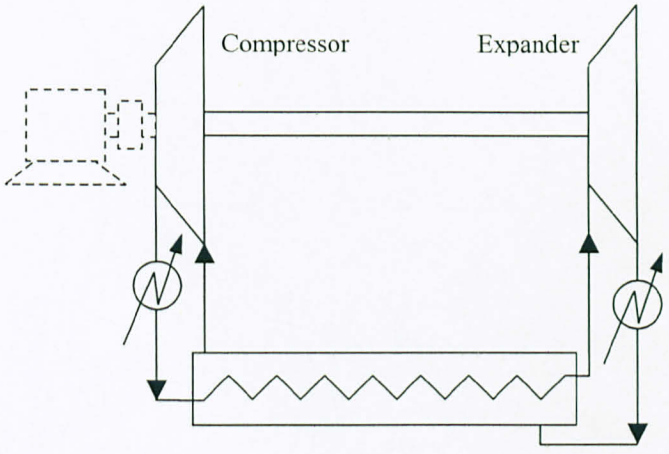


Figure 3.7: Expander/ Compressor as a refrigeration system (Ralph James *et al.*).

After the gas expands through the turbine (Figure 3.7), the exhaust temperature will be very low and capable of absorbing significant amounts of energy at low temperature in a

heat exchanger. Since the gas discharged from the heat exchanger will be at low temperature, a regenerative heat exchanger can be added to return the gas to its original state and simultaneously provide additional cooling for the compressed gas prior to entry into the turboexpander. Lower turboexpander inlet temperatures will result in lower exhaust temperatures and greater refrigeration capacity. The compressor will take in gas at low pressure and at a temperature depressed below that of ambient air, e.g., at a pressure of 1atm and a temperature of 273K. During compression the pressure and temperature of the gas will be increased then the heat exchanger is used to cool the air to above ambient air temperature. The compressor work done is shared by the gas turbine and an external power source, both of which are connected to the compressor mechanically (Ralph James *et al.*).

Turbo-expanders are very widely used as sources of refrigeration in industrial processes such as: the extraction of ethane as well as natural gas liquids (NGLs) from natural gas; the liquefaction of gases and other low-temperature processes. Turboexpanders have been in wide use since about 1950 for the separation of gases by partial condensation and the number of applications is rapidly increasing (Ralph James *et al.*).

3.4. Ammonia Vapor Absorption Chiller System

Absorption chillers use heat instead of mechanical energy to provide cooling. There are two types of absorption chillers based on the source of heat supplied to the system. "Indirect-fired" absorption chillers use steam, hot water or hot gases steam from a boiler, turbine or engine generator, or fuel cell as their primary power input. These chillers can be well suited for integration into a combined heat and power (CHP) system for buildings by utilizing the rejected heat from the electric generation process, thereby providing high operating efficiencies through use of otherwise wasted energy. "Direct-fired" systems contain natural gas burners; rejected heat from these chillers can be used to regenerate desiccant dehumidifiers or provide hot water (Guha Industries).

A thermal compressor consists of an absorber, a generator, a pump, and a throttling device. The two most common refrigerant/ absorbent mixtures used in absorption chillers are water/lithium bromide and ammonia/water. Commercially absorption chillers can be single-effect or multiple-effect.

3.4.1. Single-effect Absorption Chillers

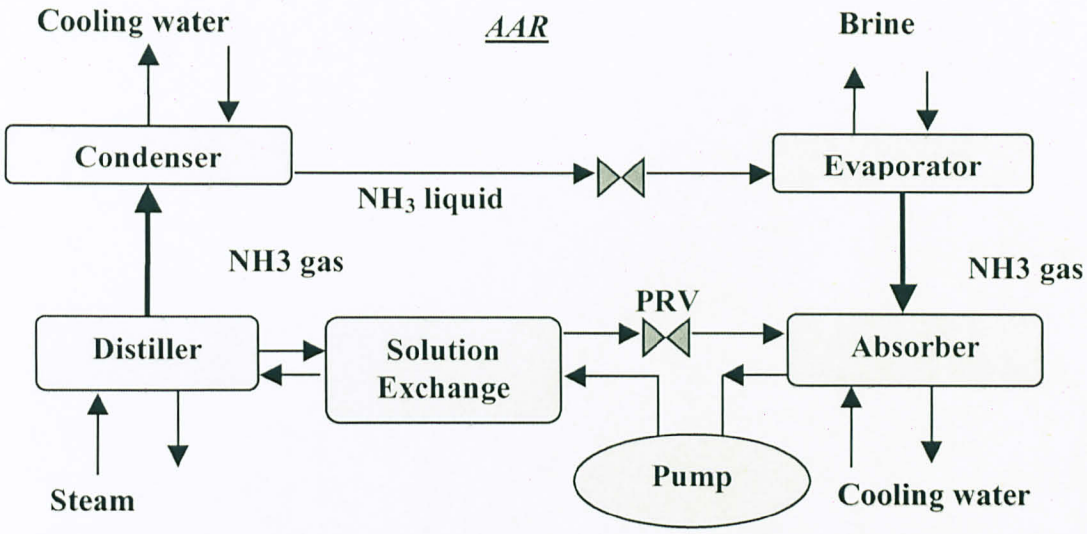


Figure 3.8: Single-effect absorption chiller cycle (Guha Industries)

The basic absorption cycle employs two fluids, the absorbate or refrigerant, and the absorbent. In Figure 3.8, water is used as the refrigerant and ammonia as the absorbent. These fluids are separated and recombined in the absorption cycle. In the absorption cycle the low-pressure refrigerant vapor is absorbed into the absorbent releasing a large amount of heat. The liquid refrigerant/absorbent solution is pumped to a high-operating pressure generator, which requires significantly less electricity than that for compressing the refrigerant of an electric chiller. Heat is added at the high-pressure generator from a gas burner, steam, hot water or hot gases. The added heat causes the refrigerant to desorb from the absorbent and vaporize. The vapors flow to a condenser, where heat is rejected and condense to a high-pressure liquid. The liquid is then throttled though an expansion valve to the lower pressure in the evaporator where it evaporates by absorbing heat and thus providing useful cooling. The remaining liquid absorbent in the generator passes through a valve, where its pressure is reduced, and then is recombined with the low-pressure refrigerant vapors returning from the evaporator so the cycle can be repeated. (Guha Industries)

3.4.2. Multiple-Effect Absorption Chillers

Multiple-effect absorption chillers are more efficient. In a single-effect absorption chiller, the heat released during the chemical process of absorbing refrigerant vapor into the liquid stream, rich in absorbent, is rejected to the environment. In a multiple-effect absorption chiller, some of this energy is used as the driving force to generate more refrigerant vapor. The more vapor generated per unit of heat or fuel input, the greater the cooling capacity and the higher the overall operating efficiency.

A double-effect chiller uses two generators paired with a single condenser, absorber, and evaporator. It requires a higher temperature heat input to operate compared to single effect absorption chiller and therefore they are limited in the type of electrical generation equipment they can be paired with when used in a CHP System.

Compared with mechanical chillers, absorption chillers have a low coefficient of performance (COP = chiller load/heat input). However, absorption chillers can substantially reduce operating costs because they are powered by low-grade waste heat. Vapor compression chillers, in contrast, must be motor or engine driven.

A single-effect absorption machine means all condensing heat cools and condenses in the condenser. From there it is released to the cooling water. A double-effect machine adopts a higher heat efficiency of condensation and divides the generator into a high-temperature and a low-temperature generator.

Absorption cooling may be worth considering if the process requires cooling, and if at least one of the following applies:

- The plant has a CHP unit and cannot use all of the available heat, or if considering a new CHP plant
- Waste heat is available
- A low-cost source of fuel is available
- The boiler efficiency is low due to a poor load factor
- The plant site has an electrical load limit that will be expensive to upgrade
- The plant site needs more cooling, but has an electrical load limitation that is expensive to overcome, and yet has an adequate supply of heat.

In general, absorption cooling is suitable when a source of free or low-cost heat is available, or if objections exist to using conventional refrigeration. Essentially, the low-cost heat source displaces higher-cost electricity in a conventional chiller.

3.4.3. Comparison of refrigeration systems

Electrical energy, and hence the operating cost of compressors, is becoming expensive everyday. LiBr (Lithium Bromide) based absorption units have an inherent temperature limitation of $+7^{\circ}C$. Wherever there is cheap heat energy available, ammonia absorption refrigeration will be the ideal choice. The following table provides a quick guide to evaluate which system would work to the plant advantage (Guha Industries, 2007)

Table 3.1: Comparison of Refrigeration systems (Guha Industries, 2007)

	Compression	**Ammonia Absorption**	**Lithium Bromide**
Electrical Energy	Runs fully on electrical power. Varies from 1KW/TR at 0°C to 2KW/RT at -25°C	Typically uses about 7%-10% of compressor electric energy for pump.	Typically uses about 5% of compressor electric energy for pump.
Refrigerant used	Varies with application	Ammonia, with water as absorbent. Environment friendly, and low cost refrigerant	Water, with Lithium Bromide as absorbent. Absorbent charging / disposal is expensive
Temperature limitation	Depends on refrigerant	Up to -33°C and lower.	Only up to +7°C
Thermal energy required	None	Low pressure steam, hot water or equivalent waste heat sources	Low pressure steam (single stage units) and high pressure steam (two stage units)
Thermal energy quantity (as steam)	None	Varies from about 9kg/TR (at 0°C operation) to 11kg/TR (-25°C operation)	About 8kg/TR (single stage. About 5kg/TR (two stage) for +7°C operation
Installation	Indoor installations	Skid-mounted outdoor installations. No building cost	Skid-mounted indoor / outdoor installation
Remarks	Requires standby for critical service. Vulnerable to fluctuations in cooling water temperature. Moving parts wear and tear means low operating life	Low maintenance, industrial standard heat-exchangers as per TEMA. Stand-by pump for reliability. Flexible operation. Higher steam temperatures can offset cooling water temperature fluctuation. Operation is easily understood and picked up by ammonia compressor operators	Vulnerable to fluctuations in cooling water temperature. Vacuum service leakages are difficult to identify. Refrigerant may need to be changed if exposed to air during maintenance. Exotic material construction. Site repair and maintenance is difficult

Chapter 4

RESEARCH METHODOLOGY

The steps taken for this research are elaborated as follows.

4.1. Data collection at Gas Plant X

The data is taken on a particular date of September 27th 2005, mostly on feed composition, product specifications, temperature, pressure and flowrate of all the main streams. The date of collecting data is based on the date of best performance of the plant, when oscillation of the plant operation is less. Data collection is done on the PI system, using computer system to extract data from control room.

4.2. Data validation

Data validation is the comparison of actual data obtained from the plant with the ICON simulation model, in which ICON is PETRONAS in-house process simulator. There are two sets of comparison performed i.e. base case comparison and modified case comparison. Base case comparison means the comparison of data collected from PI system with data from original ICON simulation; while modified case comparison is done by feeding the feed conditions (composition, temperature, pressure and flowrate) from the plant control system to the ICON simulation, letting the simulation to converge, and the comparing PI data with results from the converged ICON simulation. The range of variation of ±10% is given for data validation. The purpose of data validation is to ensure ICON simulation model corresponds with actual plant performance so that the simulation model can be used for further analysis.

4.3. Pinch analysis

Relevant stream data, such as temperature and stream heat load from simulation runs, were extracted for pinch analysis. The process area is divided into two sections, which are low temperature separation area and product recovery area. For each area, problem table, composite curve, grand composite curve are produced to identify the opportunities for improvement.

4.4. Identification of energy saving opportunities

Composite curves and grand composite curves are used to identify the scope of energy improvement. In addition, design of heat exchanger network is performed based on the pinch design method discussed in previous chapter.

4.5. Generation of simulation model based on new gas processing arrangement

Based on the design of heat exchanger network generated, simulation is carried out to generate the new gas processing plant arrangement. The simulation is first done based on separate operating heat exchanger to achieve the product specification required. The change in operating conditions is carried out in order to achieve the same product specification at lower energy consumption and higher power generated from the expander. For low temperature separation area, design of cold box arrangement is done to maintain the stream temperature required and at the same time avoid temperature cross as well as minimize the cold utility consumption. For product recovery area, thermal coupling arrangement is applied for depropaniser and debutanizer. After the heat integration within each area is done, a closer look is taken to identify opportunities for heat integration across two processing areas.

4.6. Economic evaluation and potential issues for each point identified

Economic calculation is done for each area to determine the savings generated from the reduction of energy consumption and generation of power from turbo-expander.

Chapter 5

RESULTS AND DISCUSSIONS

5.1. Assumptions

Few assumptions are made for further analysis, as follows

i. Performance of pretreatment unit (PTU) and acid gas removal unit (AGRU) are assumed to be in steady state. This means the feed gas is treated to remove all contaminant to the required level of downstream process.
ii. Performance of condensate stripper is assumed to be "normal" i.e. it is able to give the product specification as required by the downstream processing unit.
iii. Utility price (HP steam, electricity and CW) is based on in-house pricing
iv. Steam price (MP, LP & LLP) is calculated based on given HP steam price
v. Operating cost for refrigeration system is calculated based on fuel and CW cost (Appendix 9)

5.2. Data validation

A particular date of best performance was chosen to collect the operating data. The results of data validation are available in Appendix 2. Observations on the model validation are given as follows:

- Some operating parameters are out of the range of ±10% variation, which is a typical plant acceptance tolerance for process data reconciliation. The reasons could be one of the following:
  - ✓ The initial feed gas to the process is different from the original model. The original feed in ICON simulation does not have benzene, toluene and xylene, but it covers up to C_{14} instead. Therefore, the performance of the process would expect some differences from the original model.
  - ✓ Metering issues could cause some discrepancies because the plant is relatively old and no calibration was done before extracting the data. Some of the instruments were not working on that particular date; therefore no data was collected.
- All pressure parameters are within the range of ±10% variation.
- Most temperature parameters are within the range of allowable variation. However some are still out of the range. It could be attributed to the reasons as stated above.

In conclusion, the validated data is deemed acceptable for subsequent analysis as the data reconciliation is within the recommended limits of ±10%. Furthermore, absolute accuracy of the data is not required during pinch analysis as what is important is the data consistency throughout the analysis.

5.3. Separation sequencing

Natural gas feed after treatment consists of many components such as methane, ethane, propane etc, which is shown in Table 5.1.

Table 5.1: Feed gas composition

	Mole fraction (%)	mole flow (kmol/h)
Methane	0.82760	12985
Ethane	0.08875	1392.0
Propane	0.04264	669.00
n-butane	0.00968	152.00
isobutane	0.00876	137.00
isopentane	0.00677	106.00
n-pentane	0.00331	52.000
n-hexane	0.00161	25.000
n-heptane	0.00093	15.000
n-octane	0.00015	2.0000
n-nonane	0.00003	0.0000
n-decane	0.00001	0.0000
n-undecane	0.00000	0.0000
n-dodecane	0.00000	0.0000
n-tridecane	0.00000	0.0000
n-tetradecane	0.00000	0.0000
Carbon dioxide	0.00209	33.000
Nitrogen	0.00767	120.00
Total		**15690**

Total feed flow fed to the process is 15690kmol/h (equivalent to 4.411kmol/s), which mainly contains methane (82%). The gas is considered as rich gas due to the concentration of propane in the feed is more than 3% mole fraction.

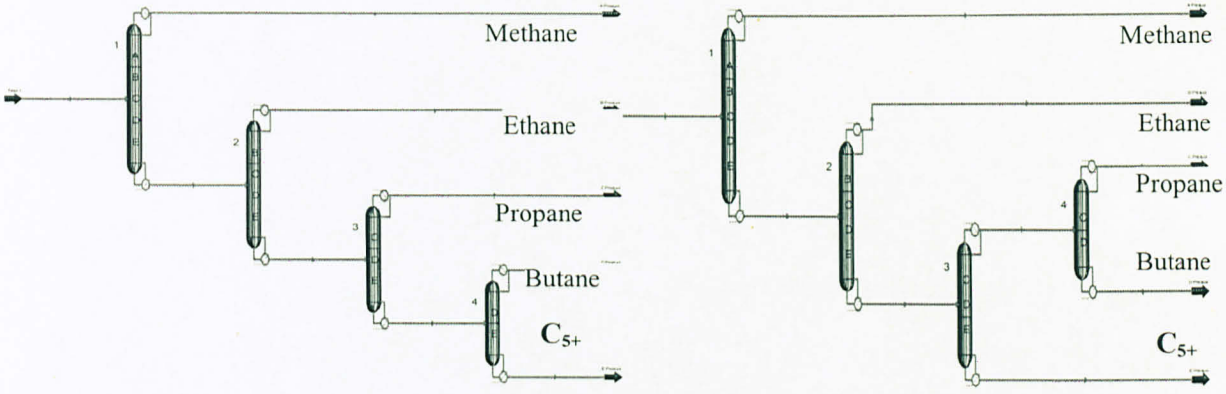
For the best result, the separation sequence needs to be determined for the best separation with minimum vapour load. This will enhance the energy consumption for reboiler and condenser of distillation column.

COLOM is a program developed by the center of Process Integration, The University of Manchester, for analyzing a variety of separation problems including column sequencing. For the sake of simplicity, we consider the separation between methane, ethane, propane, butane and C_{5+} components. The feed composition can be simplified as in the Table 5.2.

Table 5.2: Simplified feed gas composition

	Mole fraction (%)	Mole flow (kmol/s)
Methane	0.820751	3.62042
Ethane	0.094465	0.41670
Propane	0.043272	0.19088
Butane	0.018503	0.08162
Pentane	0.012755	0.05626
CO_2	0.002646	0.01167
N_2	0.007608	0.03356
TOTAL		**4.41111**

The feed condition is input to COLOM software with temperature of 25°C and pressure of 66 bars. The main objective of minimum vapor load is set in the column design. Comparison is done between Indirect Sequence design (sequence 1) and Direct Sequence (sequence 2). Direct sequence is to separate methane, ethane, propane, butane and C_{5+} components sequentially (Figure 5.1a). Indirect sequence is to separate methane, ethane, C_{5+} components, propane and butane sequentially (Figure 5.1b).



a. Direct sequence b. Indirect sequence

Figure 5.1: Direct Sequence and Indirect Sequence

These are two sequences that give the minimum total vapor load in separation of given mixtures. Direct sequence gives 8.4932kmol/s while indirect sequence gives

8.2965kmol/s. In direct sequence, column 3 is to separate between propane and C_{4+} components while column 4 is to separate between butane and condensate (C_{5+}). In indirect sequence, the sequence of separation is in the opposite way; column 3 to separate propane and butane from condensate, while column 4 is to separate between propane and butane. Therefore, for comparison purpose, the third column of direct sequence will be compared with the forth column of indirect sequence and vice versa, which is shown in Table 5.3. The first and second column of both sequences have the same function, which is to separate methane and ethane product.

Table 5.3: Comparison of direct and indirect sequences from COLOM program

Design internal details		Direct Sequence	Indirect Sequence
Column 3	Above feed vapor flow (kmol/s)	2.236	2.234
	Above feed liquid flow (kmol/s)	2.050	2.048
	Below feed vapor flow (kmol/s)	2.236	2.234
	Below feed liquid flow (kmol/s)	2.415	2.314
	Above feed minimum reflux ratio	10.03	10.02
	Above feed actual reflux ratio	11.03	11.02
	Below feed minimum reflux ratio	11.99	11.45
	Below feed actual reflux ratio	12.99	12.45
	Number of trays	11	27
Column 4	Above feed vapor flow (kmol/s)	0.095	0.294
	Above feed liquid flow (kmol/s)	0.015	0.028
	Below feed vapor flow (kmol/s)	0.095	0.294
	Below feed liquid flow (kmol/s)	0.194	0.393
	Above feed minimum reflux ratio	0.175	0.098
	Above feed actual reflux ratio	0.192	0.107
	Below feed minimum reflux ratio	2.40	1.47
	Below feed actual reflux ratio	2.42	1.48
	Number of trays	28	33

Although direct sequence has slightly more total vapor load, it is more favorable compared to indirect sequence as less number of trays are required for the third and

fourth column as well as in the separation of condensates, vapor and liquid load for both above and below feed. Moreover, if the distillation columns have both reboiling and condensation supplied by utilities, then the direct sequence often requires less energy than the indirect sequence. This is because the light material is only vaporized once in the direct sequence. From the composition of the feed gas, composition of propane is much more than butane and C_{5+} components; therefore, avoiding vaporization of propane twice will help to reduce the required energy consumption. Therefore, direct sequence is more favorable in most gas processing plant.

5.4. Pinch analysis

The studied gas processing plant was commissioned in early 1990's adopting the industry standard single-stage (ISS) NGL recovery plant. The plant was originally designed to provide a net residue gas product flowrate of 250 MMSCFD at an ethane recovery level of 80%. However, the recovery of 80% had never been sustained since commissioning. Recovery levels ranged from 65 to 72%, and attempts to increase recovery by changing operating conditions resulted in plant instabilities and upsets (Adam *et al*). During the Asian economic crisis of 1997-1998, ethane and its derivative (ethylene) enjoyed tremendous demand and high pricing, i.e. US$350/tonne in 1996 and US$550/tonne in 1998. In 2003, the plant has done some retrofits to switch from ISS to Gas Subcooled Process (GSP) operating conditions in order to adapt to the increase in demand for ethane and to improve the plant operating stability. However, as discussed earlier, energy efficiency has become the main concern of all plants' operation and gas processing plant is no exception. A typical gas processing plant consists of cryogenic section and recovery section. Cryogenic section consumes a huge amount of energy, especially by its cold utility due to the sub-ambient operating condition. It requires high amount of compression to meet its cooling requirement to a very low temperature. At the recovery section, steam consumption for reboiling purpose contributes significantly to its utility cost. Therefore utility cost becomes the major concern of most natural gas processors in order to improve the plant profitability. The scope of this research is focus on the low temperature separation unit and product recovery unit, which are the main processing area consuming huge amount of heating and cooling utilities.

5.4.1. Low temperature separation unit

Low temperature separation unit (LTSU) is a cryogenic section, which operates at sub-ambient condition. Its main purpose is to separate ethane-plus components from sales gas. A typical LTSU consists of turbo-expander, cold box and demethanizer column as shown in Figure 5.2.

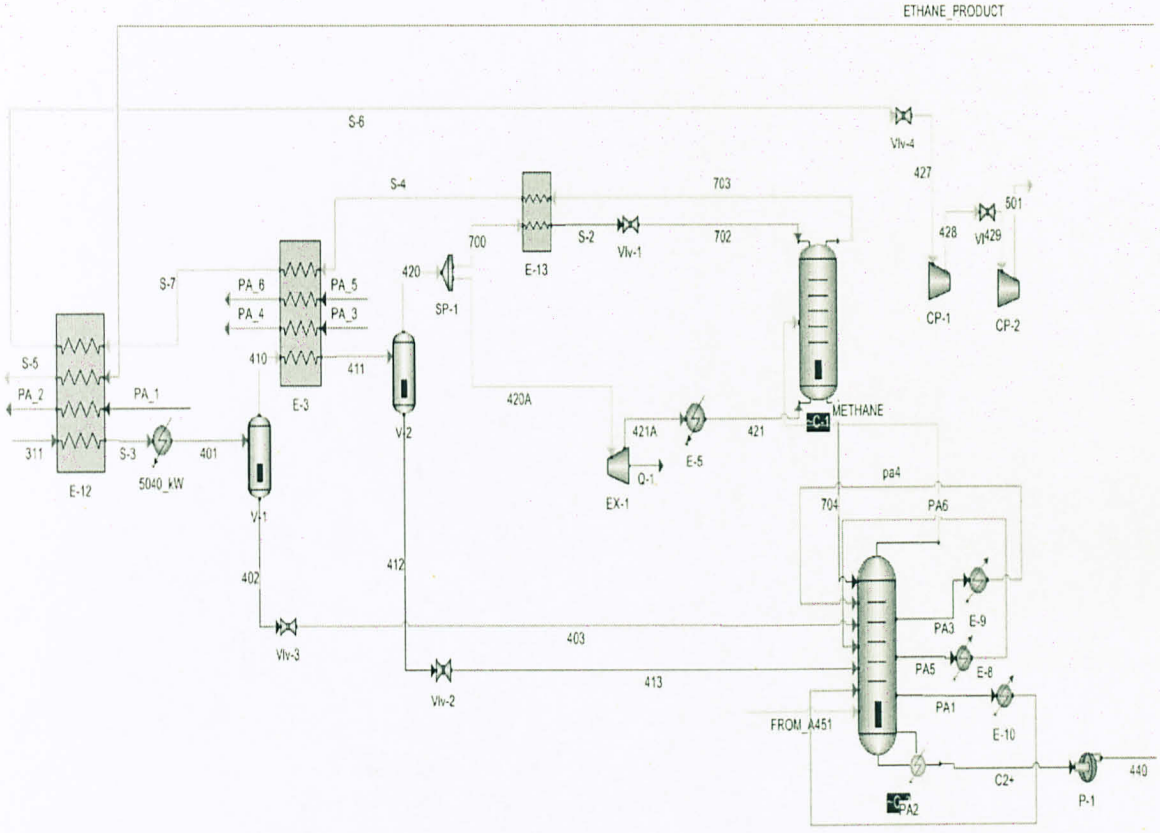


Figure 5.2: LTSU flowsheet – Base case

Data extraction is done for all the main streams, which mainly focuses on stream temperature and heat load, as shown in Table 5.4.

Table 5.4: Stream table for LTSU area

Stream name		TS [C]	TT [C]	ΔH [kW]	CP [kW/C]
1 [311]	Hot	24.80	-36.50	-20123.30	328.27
2 [410 TO 411]	Hot	-36.50	-58.50	-7590.00	345.00
3 [700 to S-2]	Hot	-58.50	-98.00	-6383.05	161.60
4 [703 to S-6]	Cold	-100.80	20.20	18492.80	152.83
5 [ETHANE_PRODUCT]	Cold	-27.70	21.80	4165.28	84.15
6 [PA1 to PA2]	Cold	0.18	8.24	2778.00	344.75
7 [PA5 to PA6]	Cold	-90.47	-85.00	1186.67	217.06
8 [PA3 to PA4]	Cold	-77.31	-60.04	2444.01	141.51

From the stream data in Table 5.4, if there is completely no heat integration in the area, the cold utility needed is the summation of all hot streams, i.e. [feed 311], stream [410 to 411] and stream [700 to S-2], which is equivalent to 34096.35 kW. The hot utility required is the summation of all cold streams, i.e. from stream 4 to stream 8, which is equivalent to 29063.76 kW. However, the arrangement of cold box in LTSU area allows the heat recovery from hot streams to cold streams. The composite curve is produced to identify the minimum heating and cooling requirement in the area as shown in Figure 5.3.

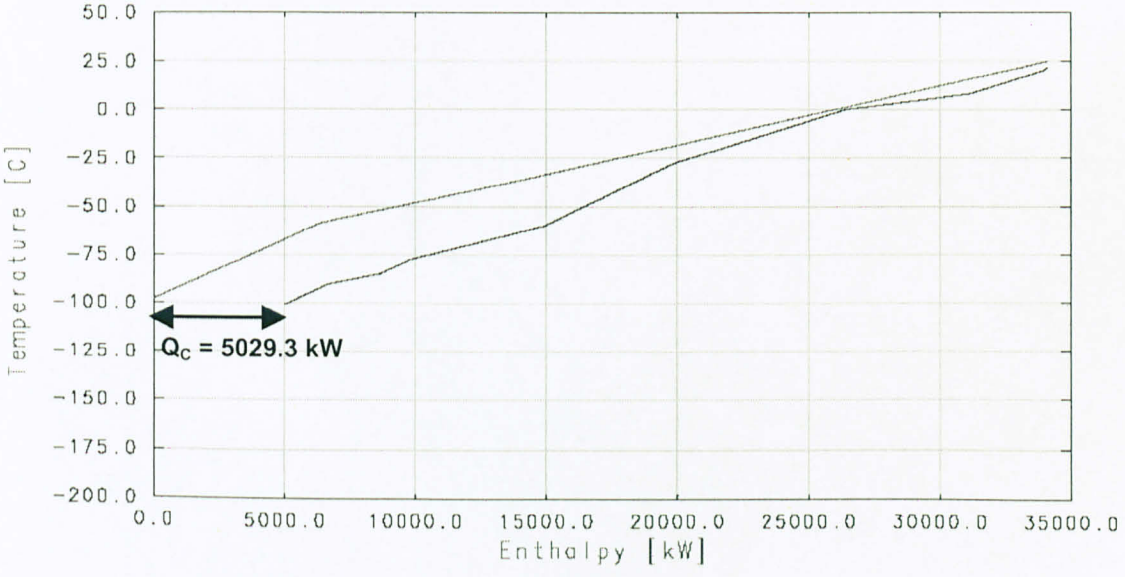


Figure 5.3: Composite curve of LTSU

The ΔT_{min} = 0.3°C is set for a typical gas phase heat recovery. From the composite curve, there is no hot utility required for LTSU but cold utility load of 5029.3 kW is required. In reality, this amount is hardly achieved due to many constrains in design. The simulation model gives the required cooling load of 5040 kW, which is quite close to the energy target. This shows the LTSU is quite well heat integrated. However, the cold utility required for this area is the main concern for all the gas processors. The temperature level for the heat exchanger is to cool from -23.3°C to -36.5°C, which requires low temperature refrigeration. Therefore the operating cost to supply refrigeration at this temperature level is high. The target is to eliminate the cooling load or at least to reduce the cooling load in this LTSU area to as low as possible. Another concern in this area is the power generated by the turbo-expander EX-1, which could be used for compression of sales gas to meet the customer's requirement (CP-1). Currently, the turbo-expander EX-1 could generate 1389kW, which is used to compress sales gas from 2091kPa to 2358kPa. This contributes to savings in operating cost of compressor. Therefore, the more power generated, the more savings in operating cost is realized.

Few efforts have been done to improve the energy consumption and the power generated from the turbo-expander. The first attempt is exploring the turbo expander EX-1 to extract as much power as possible from the expansion of feed gas. Originally, the pressure drop across the turbo-expander EX-1 is 3MPa, which generates cooling effect from -58.5°C to about -89°C (Fig 5.4). In order to maintain the same pressure drop and enhance the power generated, the inlet stream to the turbo expander EX-1 needs to be heated to higher temperature. Theoretically, hotter stream will generate more power at the same pressure drop across the turbo expander. Therefore, a heater Q4new is needed to increase the temperature from -50°C to -30°C. The heating load of Q4new needed is 2880kW. However, the hotter the inlet to turbo-expander, the hotter its outlet will be produced. The outlet of expander is acting as a reflux stream to enhance ethane recovery. The colder the expander outlet fed to the absorber, the better the ethane recovery. Therefore, in order to maintain ethane recovery, it is necessary to cool expander outlet [421A] at -68°C to the original temperature of -89°C, which requires 4660kW of cooling load (Q5new). The power produced is 1885kW, which is 36% more than the original 1389kW. More cooling and heating load required is a penalty for more power generated.

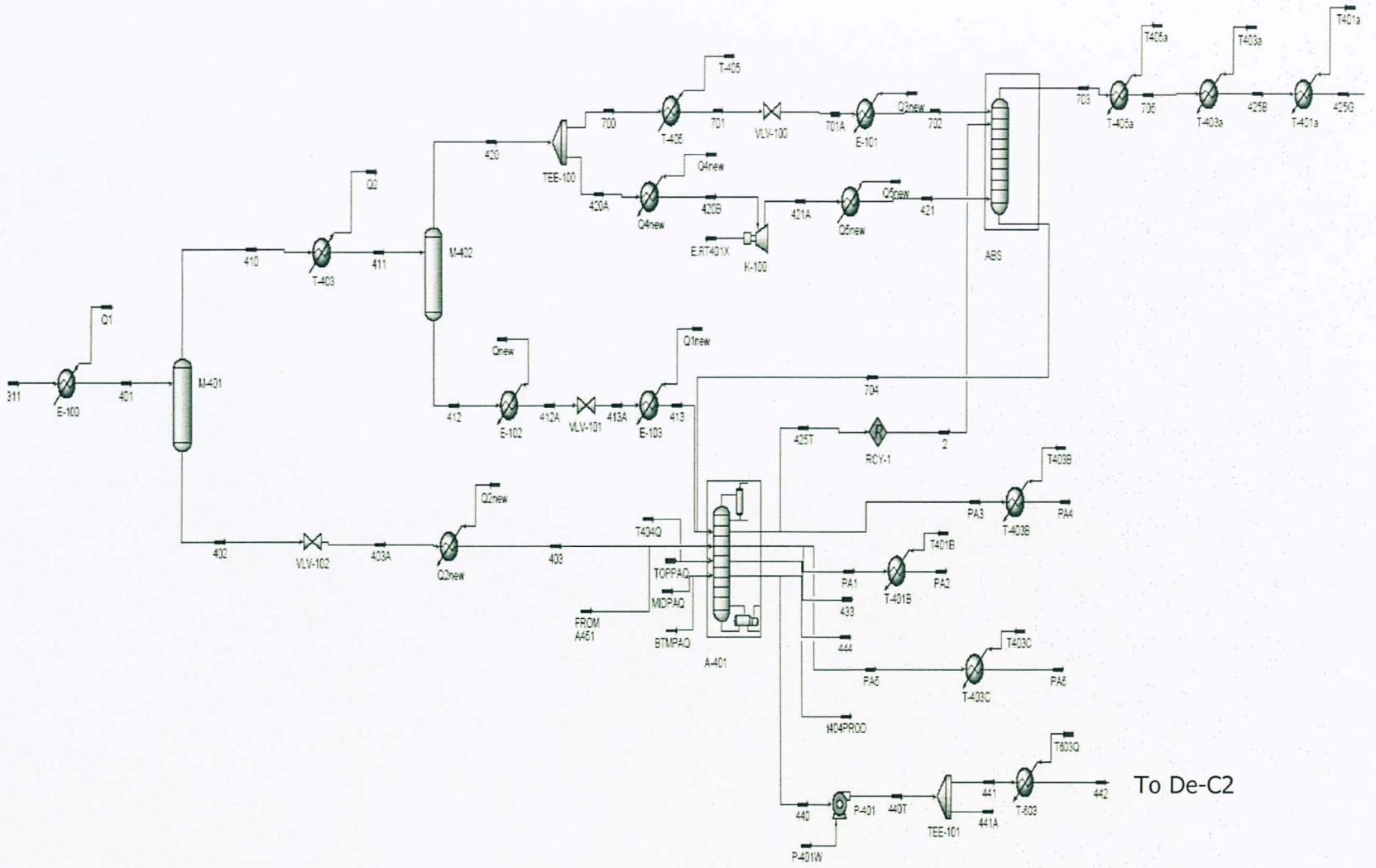


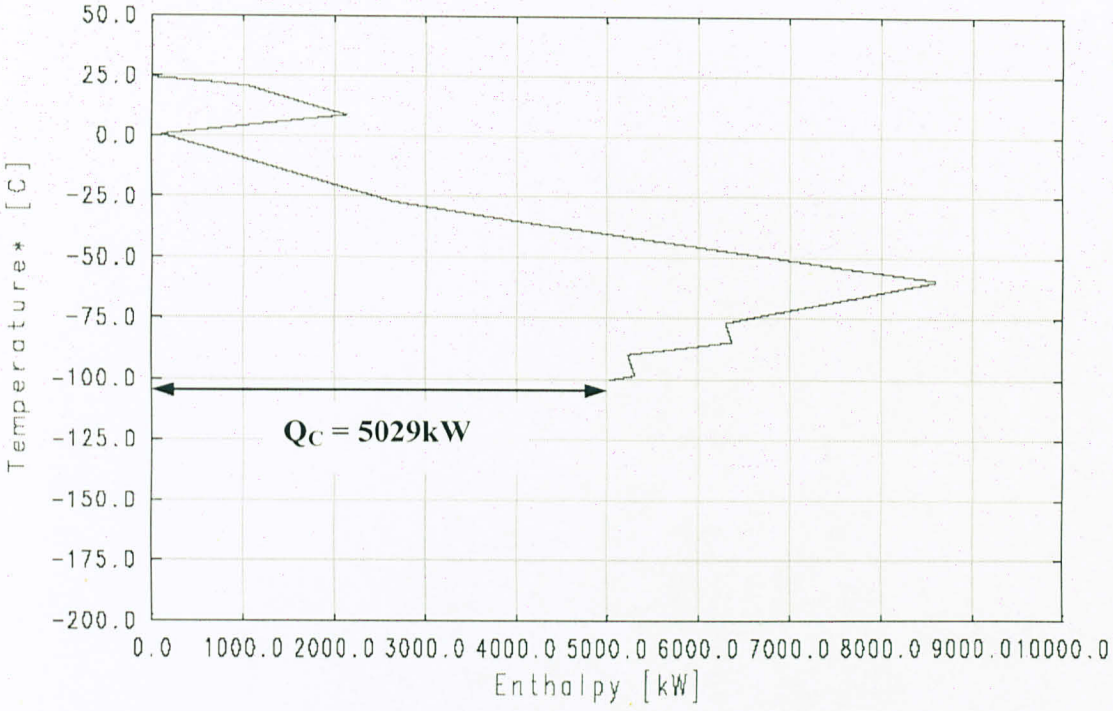
Figure 5.1: LTSU with additional heaters and coolers in separate operation

The second attempt is to have a close look at expansion valve in this area. After the phase separator M-401, the feed stream [410] is cooled down to -50°C instead of -58.5°C in the original case, which is to generate more gas phase in the second phase separator. The vapor fraction of feed stream to second phase separator M-402 is 0.95525 (12606kmol/h vapor) compared to 0.86808 (11455kmol/h vapor) in the original case. The purpose of more vapor flow in the separator is to produce more power extracted as more flow is going through the turbo expander. This also provides more reflux flow from outlet of turbo expander to enhance the ethane recovery.

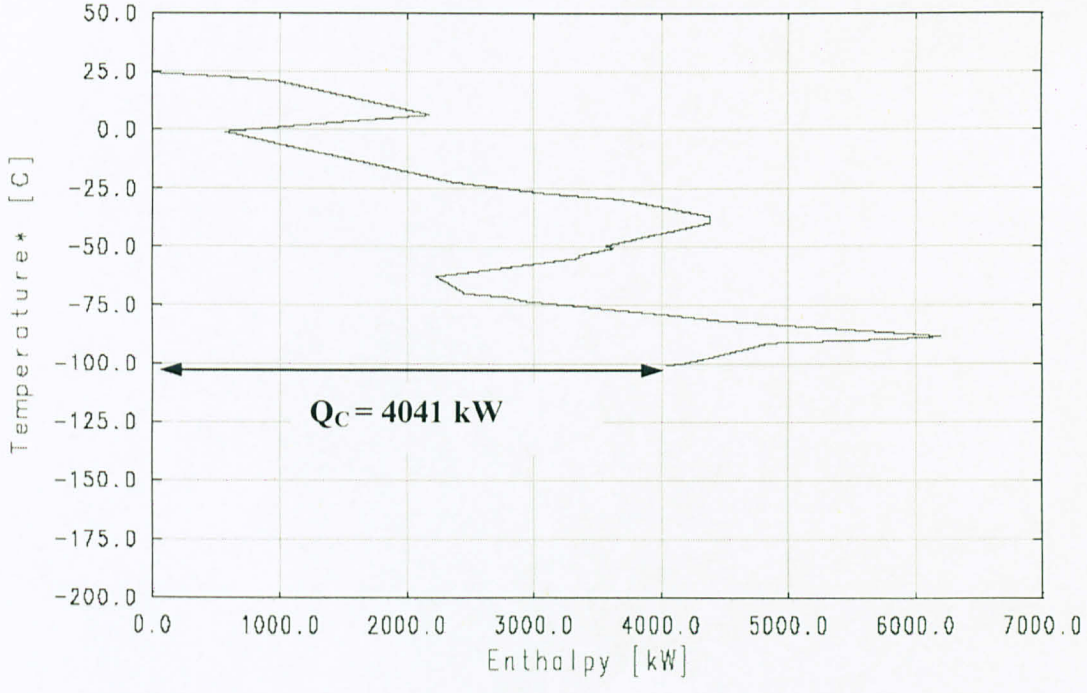
The bottom liquid [402] of the phase separator M-401 is flashed down to the demethaniser operating pressure at 2468kPa via expansion valve VLV-102. The pressure drop is 2977kPa, which results in cooling of the stream from -36.5°C to -54°C. Therefore, in order to utilize this cooling effect, a heater Q2-new is added to heat the stream [403A] from -54°C to -40°C before feeding it to the demethaniser column A-401. By doing this, a heating load is introduced below the pinch, which reduces the cooling load required. Moreover, a partially vaporized feed to demethaniser will contribute to less reboiler duty required.

The same rule applies to the bottom liquid of the second phase separator M-402. This stream is flashed down to the demethaniser operating pressure at 2575kPa via expansion valve VLV-101. The pressure drop of 2805kPa generates the cooler stream from -58.5°C to -81.5°C in the original case and -50°C to -70.8°C in the modified case. A heater E-103 is introduced to heat the stream [413A] from -70.8°C to -75°C before being fed to the column.

LTSU consists of few expansion valves to flash down stream pressure before entering the demethaniser column A-401. The pressure drop across each expansion valve is critically high. The expansion valve produces the expansion of stream without any energy recovery by the Joule-Thomson effect. Introducing heaters after the expansion valve generates more heat recovery pocket from the Joule Thompson valve. This is clearly shown in the grand composite curve (Figure 5.5).



a. GCC of original case



b. GCC of modified case

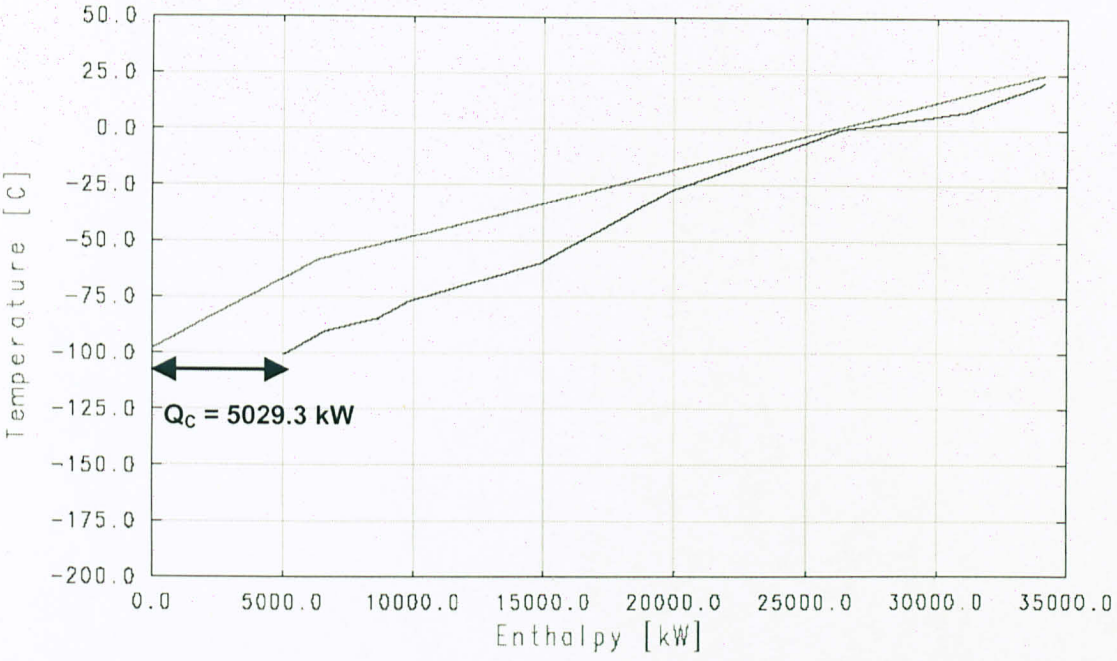
Figure 5.5: Comparison of Grand Composite Curve (GCC) of original and modified case

The new stream data is extracted for the new configuration. In data extraction, some of the streams are divided into few segments due to varied heat capacity CP. In Sprint program, CP is assumed to be constant, which leads to the inconsistency between Sprint results and simulation results. Therefore, the division of stream into smaller segments shall ensure the varied heat capacity from one condition to another is taken into account.

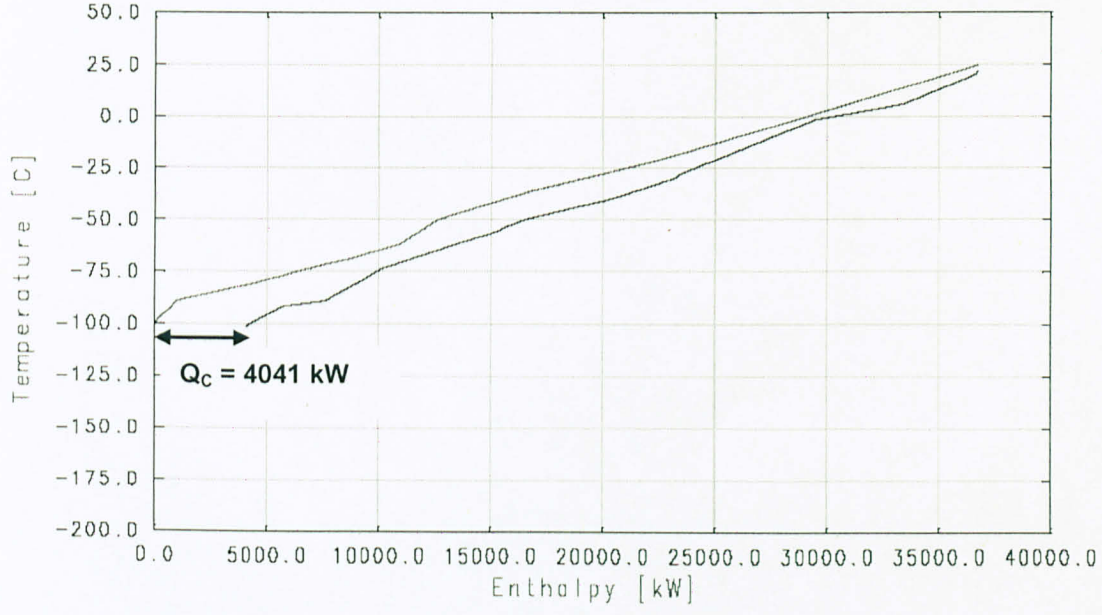
Table 5.5: Stream table for modified LTSU area

Stream name		TS [C]	TT [C]	ΔH [kW]	CP [kW/C]
1 [FEED 311]	Hot	24.80	-21.84	-14527.80	311.49
	Hot	-21.84	-36.50	-5591.70	381.40
2 [410N TO 411A]	Hot	-36.50	-50.00	-3998.00	296.15
3 [700 to S24]	Hot	-50.00	-62.00	-1714.40	142.87
	Hot	-62.00	-71.00	-2858.00	317.56
	Hot	-71.00	-87.00	-2332.00	145.80
	Hot	-87.00	-99.50	-1180.00	94.64
4 [421A_N to S12]	Hot	-69.42	-81.10	-2173.50	186.10
	Hot	-81.10	-89.00	-2230.00	282.28
5 [413A_N to 413N]	Hot	-70.78	-75.00	-105.42	25.00
6 [403N to 403]	Cold	-53.93	-40.00	1142.50	81.99
7 [703 to S11]	Cold	-101.37	-80.00	3621.81	169.48
	Cold	-80.00	-36.70	6598.40	152.30
	Cold	-36.70	20.20	8100.02	142.42
8 [ETHANE_PROD]	Cold	-27.58	21.80	4191.70	84.90
9 [420AN to 420BN]	Cold	-50.00	-30.00	2872.20	143.61
10[PA1 to PA2]	Cold	-1.47	6.00	2242.20	300.20
11[PA5 to PA6]	Cold	-91.64	-89.00	1426.00	540.36
12[PA3 to PA4]	Cold	-73.65	-56.00	2486.30	135.17

From the composite curve of the modified gas plant, more heat recovery pocket is introduced, which will generate the improvement in cold utility consumption. This reflects clearly in the composite curve.



a. Composite curve of original LTSU design



b. Composite curve of modified LTSU design

Figure 5.6: Composite curve of LTSU

Introducing heaters after expansion valves moves the hot composite curve slightly downward while cooler after turbo expander directs cold composite curve to move upward and slightly extend to the left. This generates more overlap area between hot and cold composite curves, which reflects more heat recovery. This reduces cold utility required for LTSU area, which is 4041kW compared to the original 5029kW. This is clearly seen in the composite curve by the closer gap between the hot and cold composite curves.

The cold utility is required at very low temperature, -100°C, which is very costly as high compression power is required. Therefore, in order to reduce the refrigeration load at this low temperature, a fixed cooling load is supplied at much higher temperature, say at -12°C. By supplying about 2562 kW cooling load at -12°C, the pinch point is shifted from 24.8°C to -12°C. Therefore, a heating load of 333kW is required and cooling load is reduced to 2161kW. However, the heating load is not a problem in LTSU area because it is at very low temperature, about 25°C. This load can be easily supplied by running water or even better can be heat integrated with any product rundown at downstream. The cooling load now is the summation of fixed refrigeration load of 2562kW at -12°C and another load of 2161kW, which is 4723kW. Although this load is more than the one before adding the constant cooling load, this could have cheaper operating cost due to less compression cost required by refrigeration system. A new composite curve is produced for the new arrangement.

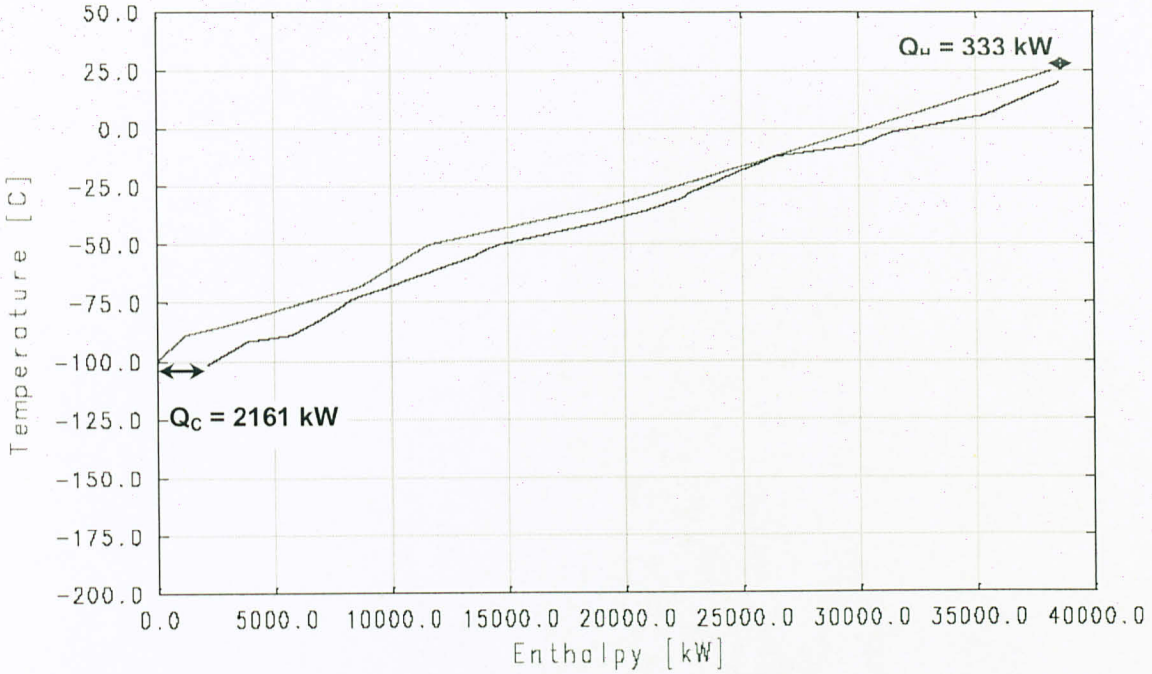


Figure 5.7: Composite curve of modified LTSU with constant cooling load

The problem table is tabulated based on the data extraction, which shows the pinch points of the area. The pinch points are at -12°C for cold stream and at -11.89°C for hot stream with ΔT_{min} = 0.105°C. This requires 333kW for hot utility and 4723kW for cold utility (including 2562kW of constant cooling load at -12°C)

The grid diagram for the modified LTSU is shown in Figure 5.8. The grid diagram is divided into two sections, which are above pinch and below pinch sections.

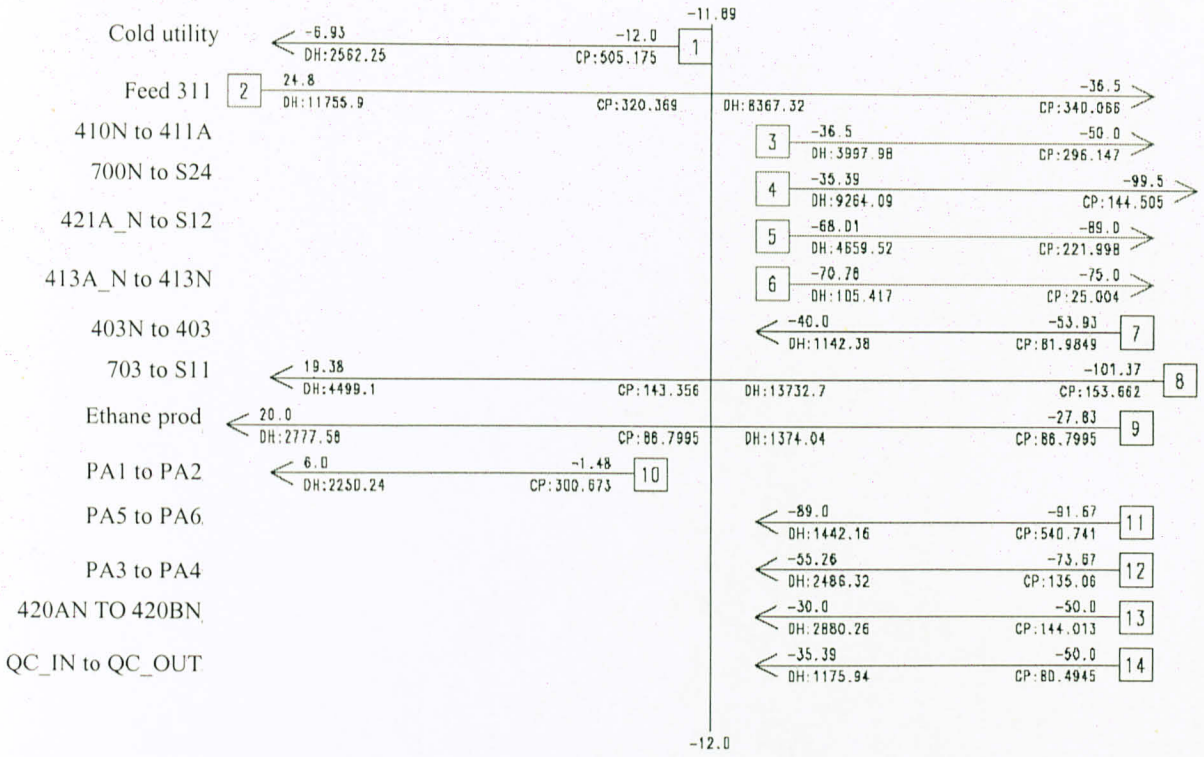


Figure 5.8: Grid Diagram of Modified LTSU

From the grid diagram, heat exchanger network is designed based on the rules of heat exchanger network design and pinch design mentioned earlier. The network data report is produced by the Sprint program shown in Appendix 9. The heat exchanger network is expressed in the form of series of cold box and heat exchanger arrangement, which is shown in Figure 5.9. In the LTSU arrangement, there are 5 cold boxes, E-4, E-5, E-12, E-13 and E-14. E-4 represents for heat exchanger 15 in the network, while E-5 for HE 9, E-12 for the combination of HE {2, 3, 4, 5}; E-13 for HE {6, 7, 15, 16, 17, 18, 19}; E-14 for {8, 9, 10}, E-4 for {11, 12}respectively. Besides, there are some heaters and coolers in this area, which are indicated by HE 1 and 13 in the network.

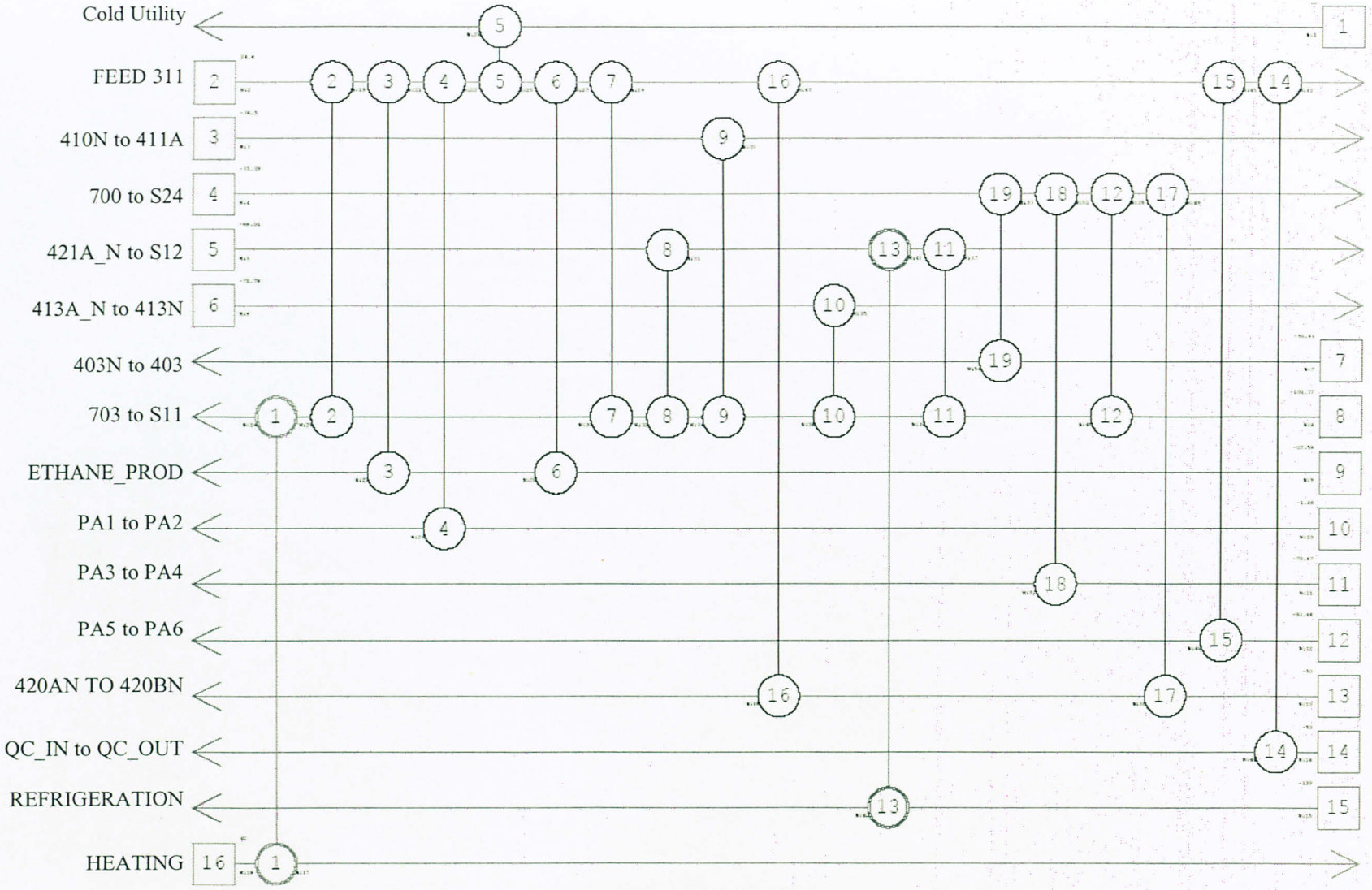


Figure 5.9: Heat Exchanger Network for new LTSU arrangement

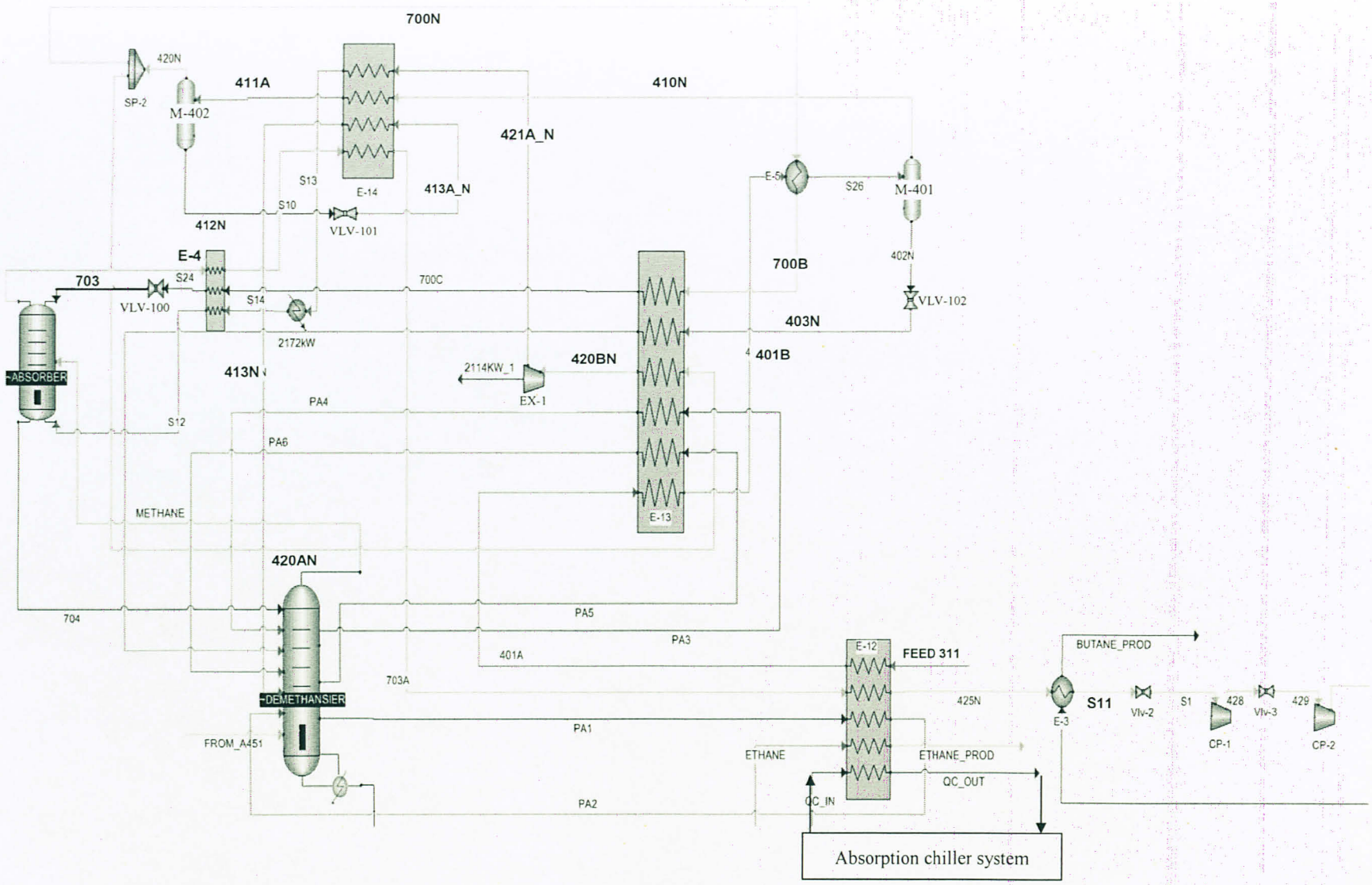


Figure 5.10: New LTSU arrangement

The cooling load in LTSU is at -85°C; instead of provide the cooling load at -100°C. This could help to reduce the operating cost for the refrigeration system.

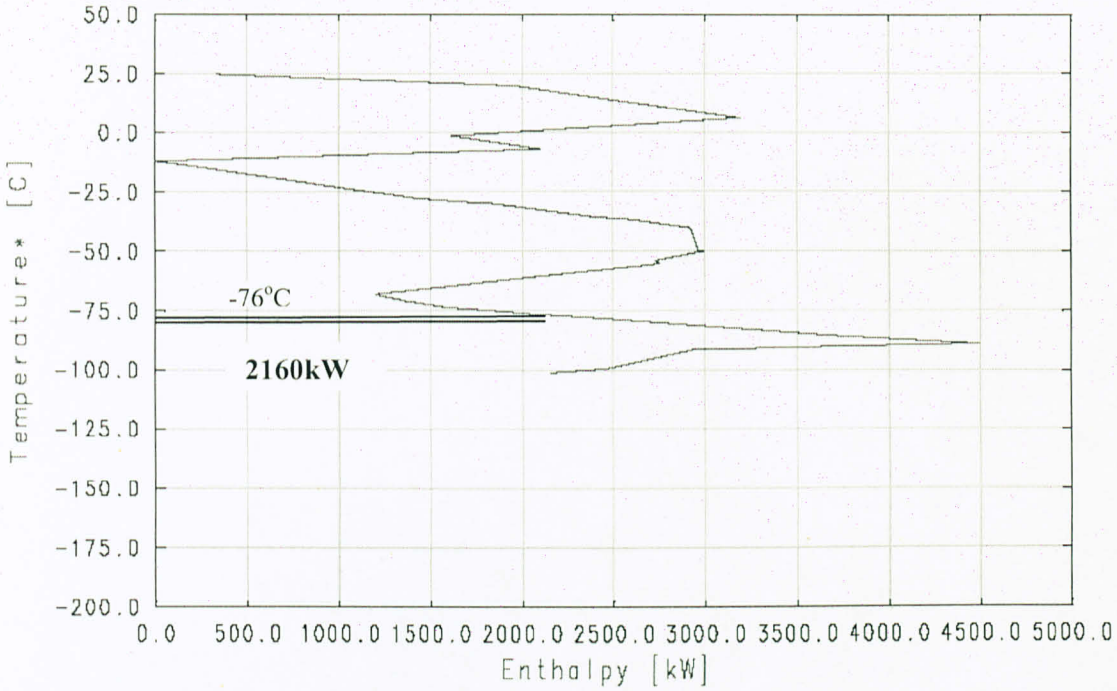


Figure 5.11: Grand Composite Curve with cooling levels

From the grand composite curve, a minimum cooling load of 2160kW is required at -76°C. However, from the network design, it requires 2172kW at -76°C and cool down to -85.5°C.

As mentioned before, a constant cooling load of 2562kW is supplied to LTSU at -12°C. This cooling load is added as a fixed quantity to the system, which is considered as a cold stream in the stream data table. This cold stream is heated from -12°C to around -7°C. The cold stream at -12°C is chosen by carrying out optimization of few temperatures around the range of -10°C. This addition of cooling load could lead to the penalty of having heating load in the system but the heating requirement is not a big issue in LTSU area due to the low temperature heating. The cold stream of -12°C heated to -7°C is mentioned as QC_IN stream in Figure 5.10. The stream is located in cold box E-12. Outlet stream named as QC_OUT, which is connected to absorption chiller systems.

The design will not be meaningful if it is not more economical than the original design. Therefore, a cost analysis should be done to ensure the new design can generate more

savings. The refrigeration cost is different at different cooling level. The colder the cooling level, the more expensive it will cost; because more compression is needed to compress the refrigerant in order to produce sufficient cooling. The detailed calculation for each cooling level is shown in Appendix 8. Cost analysis is shown in Table 5.6.

Table 5.6: Cost analysis of refrigeration in LTSU

	Old Design	New Design	
		Constant Cooling	Heat Exchanger
Temperature (°C)	-36.5	-12	-85
Cooling load (kW)	5040	2562	2172
Refrigeration used (kW)	3122	763	2684
Annual cost (RM)	1,212,840	296,431	1,042,836
Total cost (RM)	**1,212,840**	**1,339,267**	
Annual savings of new design		**-126,427**	

Table 5.7: Cost analysis of power generated and reboiler duty in LTSU

	Old design		New design	
	Load (kW)	Cost (RM)	Load (kW)	Cost (RM)
Power generated	1386	2,706,359	1885	3,680,726
Reboiler duty	4792	832,076	4487	779,116
Predicted annual savings of the new design			**RM 900,900**	

The annual cost for refrigeration used in the new design is around RM126k more expensive than the original design due to more total cooling load required. However, as mentioned earlier, the constant cooling load of 2258 kW can be easily taken care of by absorption chiller, which is much cheaper than using compression refrigeration system. Therefore, the cost of about RM 300k can be eliminated from the total cost for cooling in new design. Moreover, the new design could generate about 36% more power from the

turbo-expander. With the electricity cost of RM 226.4/MWh, additional power generated at 500kW could generate annual savings of about RM 974k for the gas plant. Therefore, the total annual savings from the new design is RM 901k inclusive of constant cooling load in refrigeration system and RM 1,197,331 if the constant cooling load is taken care of by an absorption chiller system. All savings are summarized in the Table 5.8, which constant cooling is taken care of by absorption chiller.

Table 5.8: Predicted annual savings from the new LTSU design

	Old Design	New Design	Savings
Refrigeration (RM)	1,212,840	1,042,836	170,004
Power Generated (RM)	2,706,359	3,680,726	974,367
Reboiler duty (RM)	832,076	779,116	52,960
Predicted annual savings from the new design (RM)			**1,197,331**

The new LTSU arrangement has one additional cold box (E-14) compared to the base case, which has overall UA = 2.78 x 10^6 W/K (Appendix 5). The other cold boxes in the original case can be utilized for the new arrangement as they have approximately same overall heat transfer area. The heat transfer coefficient for mixed hydrocarbons is approximately 430W/m^2.K (Gael D. Ulrich). Therefore the heat transfer area for the additional cold box is 6465m^2, which cost approximately RM 4.55 mil (according to estimated cost from Plant X). The investment cost of RM 5.23mil is the summation of the cost of new cold box and the piping modification cost, which is 15% of the new cold box cost.

Some assumptions are made for the calculation of payback period as well as the rate of return for this investment as follow:

- Interest rate: 10%
- Assumed planned life of the project is 15 years
- NPV is calculated on the basis of 15 yrs
- Payback period is calculated based on HEs cost and piping modification only
- IRR is calculated based on estimated operating cost (15% of piping modification cost) and maintenance cost (5% of piping modification cost)

The detailed calculation for net present value (NPV) and internal rate of return (IRR) is attached in Appendix 10.

$$\text{Payback Period} = \frac{\text{Total investment cost}}{\text{Total savings}} = \frac{\text{RM 5.23 mil}}{\text{RM 1.20 mil}}$$

$$= 4.5 \text{ yrs}$$

Table 5.9: Summary of economic analysis for thermal coupling arrangement

NPV	RM 1.89mil
Payback period	4.5 yrs
IRR	19%

The new LTSU has incentive for gas processors to consider. In order to give a better view of economic value of this new arrangement, more detailed economic analysis should be carried out, which is beyond the scope of this research.

5.4.2. Product Recovery Unit (PRU)

PRU area consists of three distillation columns, which are de-ethaniser, de-propaniser and debutanizer columns. This unit separates valuable liquid products in the bottom product of demethaniser such as ethane, propane, butane and condensates, which is called natural gas liquid (NGL). The flowsheet for this area is shown in Figure 5.12.

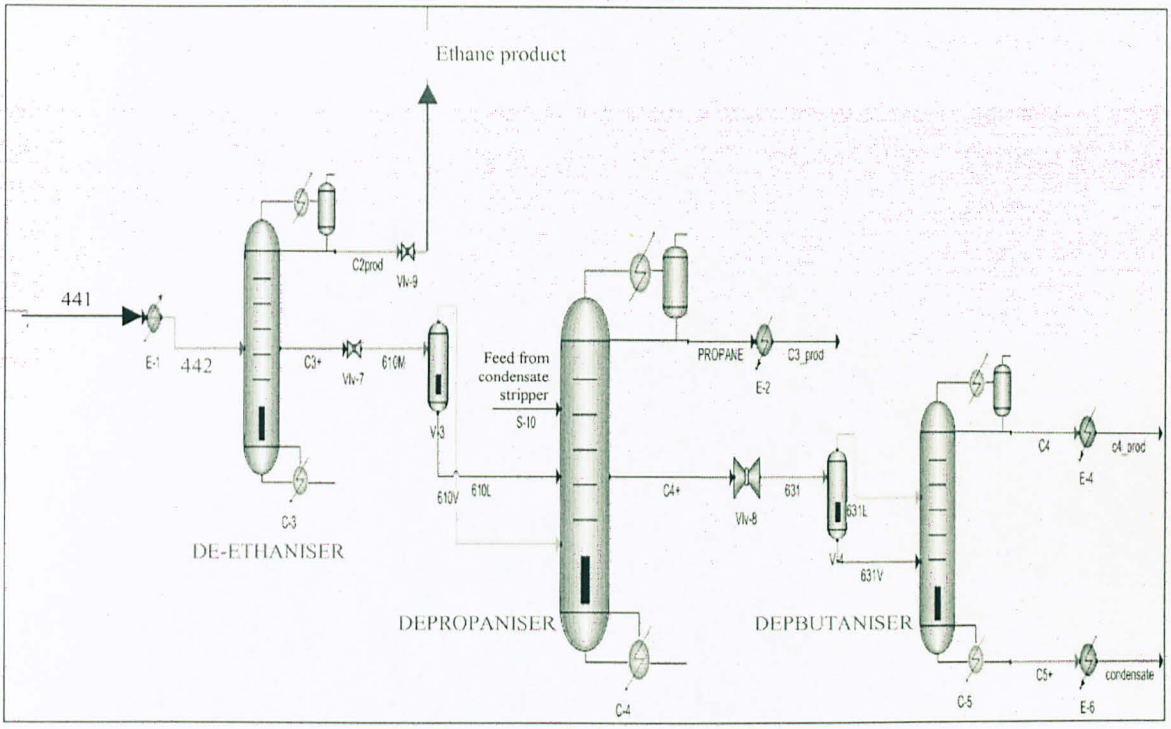


Figure 5.12: PRU flowsheet - Base case

Based on the relative volatility differences, ethane recovery is "easier" than propane recovery. However, the lower condensing temperature of ethane versus propane (-88.6°C versus -42°C, both at atmospheric pressure) is the primary reason more energy is required for ethane recovery.

Stream data extraction from simulation model is done as shown in Table 5.10.

Table 5.10: Stream table for PRU area

Stream		TS [C]	TT [C]	ΔH [kW]	CP [kW/C]
1 [ovhd C2]	Hot	6.50	3.30	-8259.01	2580.94
2 [ovhd C3]	Hot	51.00	47.00	-7575.00	1893.75
3 [C3 prod]	Hot	44.60	21.00	-609.00	25.80
4 [ovhd C4]	Hot	48.00	44.00	-4492.00	1123.00
5 [C4 prod]	Hot	44.00	20.00	-278.00	11.58
6 [C5+ prod]	Hot	118.00	30.00	-1707.75	19.41
7 [441 to 442]	Cold	25.00	37.00	1131.30	94.28
8 [btm C2]	Cold	100.00	101.00	10040.00	10040.00
9 [btm C3]	Cold	128.00	129.00	6161.00	6161.00
10 [btm C4]	Cold	118.00	119.00	3117.00	3117.00

Similarly, the cold utility requirement is the summation of all hot streams, i.e. from stream 1 to stream 6, which is 22920.75 kW without any heat recovery, and heating load required is 20449.30 kW. However, the minimum heating and cooling load is shown by conducting the composite curve as in Figure 5.13.

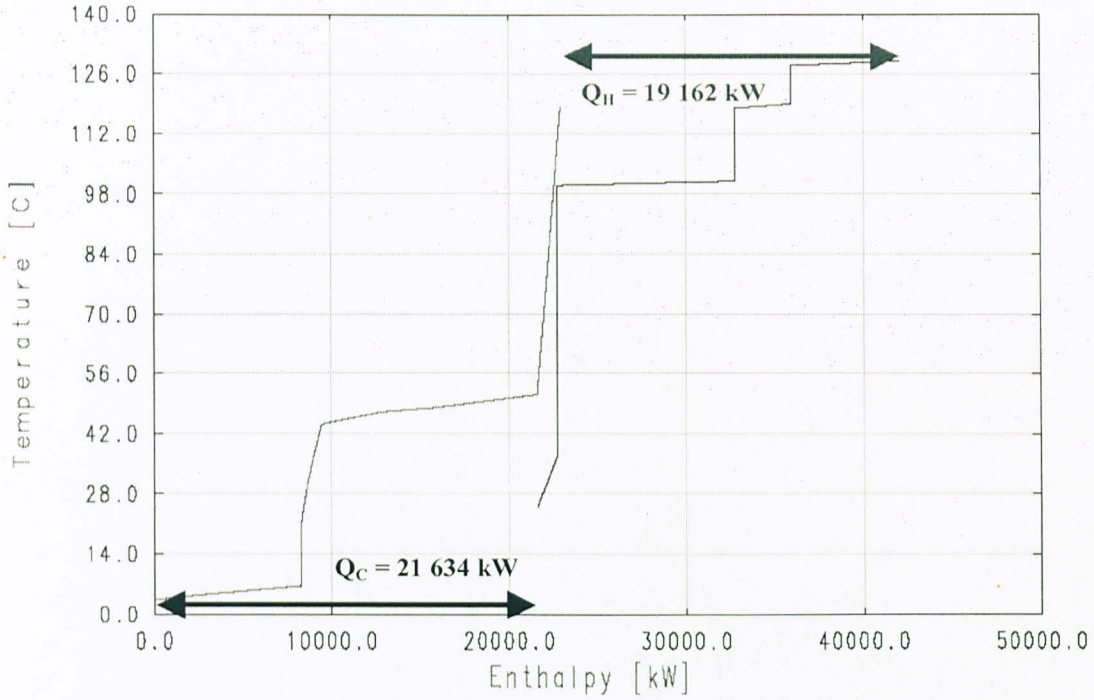


Figure 5.13: Composite Curve for PRU area

Although there is heat recovery shown in the composite curve, there is very little scope for heat recovery as the overlap between the hot and cold composite curves is very small. This limit is attributed to the pinch causing cold stream of a reboiler at about 100°C. One can reduce energy consumption by lowering the column pressure but this is not practical. Since the process does not show significant amount of overlap between hot and cold composite curves, coupled with a wide temperature span across the process, no substantial energy saving opportunities can be derived.

Thus, pinch analysis cannot make any improvement in this area due to very little scope for heat recovery; attention is given to look into the energy efficiency of the distillation processes itself. A complex column is taken into consideration to replace conventional distillation to reduce energy consumption. A prefractionator arrangement (or a thermal coupling) is introduced for propane, butane and condensates separation because the gap of boiling points between propane and butane is closer compared to that between ethane and propane.

A prefractionator arrangement consists of two columns. In the prefractionator, the lightest (propane) and heaviest components (C_{5+} components) are chosen to be the key separation. A partial condenser and a partial reboiler are applied in this arrangement to reduce the overall energy consumption. This arrangement requires 20 to 30% less energy than a conventional arrangement for the same separation duty. The reason for this difference is rooted in the fact that the prefractionator arrangement is fundamentally thermodynamically more efficient than a simple arrangement.

In a conventional arrangement, the composition of butane in the depropaniser column increases below the feed as more volatile propane decreases. However, when moving further down the column, the composition of butane decreases again as the composition of the less volatile C_{5+} components increase. Thus the composition of butane reaches a peak only to be remixed. This remixing in the conventional column is a source of inefficiency in the separation. In contrast, consider the prefractionator arrangement; a crude split is performed so that butane component is distributed between the top and bottom of the column of the prefractionator column. The upper section of the prefractionator separates propane and butane from C_{5+} components, whilst the lower section separates butane and C_{5+} components from propane. Thus, both sections remove only one component from the product of that column section and this is also true for all four sections of the main Depropaniser column. In this way, remixing effects are avoided (Robin Smith).

In prefractionator arrangement, the condenser is operating as a partial condenser with all the vapor draw is fed to the main depropaniser column. The number of trays of prefractionator and depropaniser column is maintained same as the original case for ease of comparison. Prefractionator has 39 trays with 2 feeds at 13^{th} stage and 26^{th} stage. Feed from bottom of de-ethaniser column is fed to 13^{th} stage while the feed from condensate stripper is fed to 26^{th} stage. The location of these two feeds to prefractionator is optimized to obtain the minimum energy consumption required for condenser and reboiler of prefractionator column. With this configuration, the cooling load of condenser is 2288kW at 31°C while the heating load of reboiler if 2702kW at 104°C as shown in Figure 5.14.

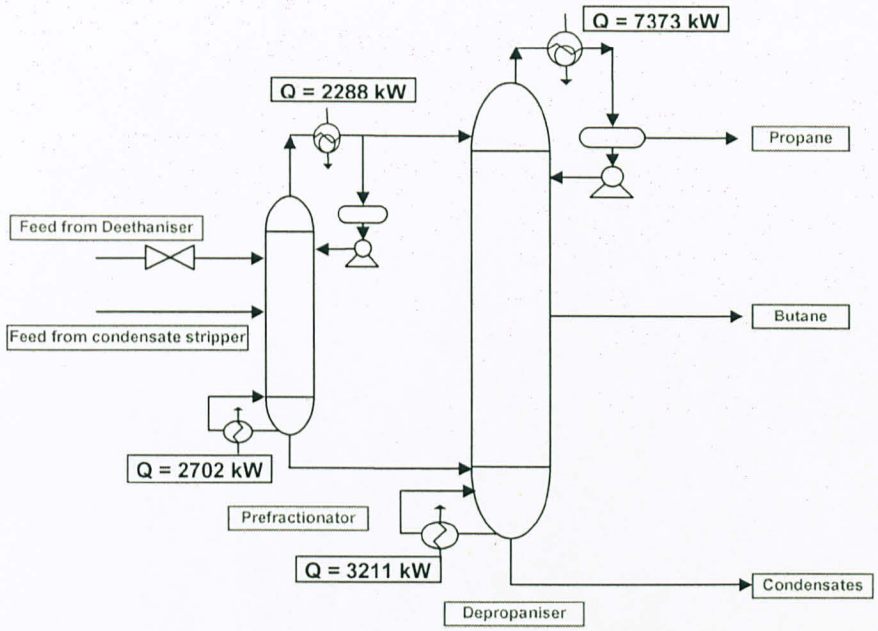


Figure 5.14: Prefractionator arrangement for depropaniser and debutanizer column

The overhead and bottom products from prefractionator are fed to the main depropaniser column at 11th stage and 28th stage. The depropaniser column operates with 36 trays. There are three products collected in this column, which are propane as overhead product, condensates as bottom product and butane as liquid side draw at 19th stage. Propane and butane products meet the purity requirement of customer at 98 mole% purity. With this configuration, it requires the condenser duty of 7373kW and 3211kW as reboiler duty. Table 5.11 shows the comparison of the new prefractionator arrangement and the conventional arrangement for depropaniser and debutanizer columns.

Table 5.11: Comparison of prefractionator and conventional arrangements

		Conventional arrangement	Prefractionator arrangement
Depropaniser	Condenser duty	9773kW @ 42°C	2288kW @ 31°C
	Reboiler duty	8630kW @ 128°C	2701kW @ 103°C
Debutaniser	Condenser duty	5635kW @ 48°C	7373 kW @ 4°C
	Reboiler duty	4313 kW @ 126°C	3211 kW @ 116°C

From the comparison, the total condenser duty of prefractionator arrangement is 9661kW compared to 15408 kW in the conventional arrangement, which is equivalent to 37% savings in terms of energy load. Similarly, prefractionator arrangement requires 5912 kW of reboiler duty compared to 12943kW of conventional arrangement, which generates 54% savings on MP steam usage. Moreover, the temperature at both reboilers in new arrangement is lower than that in original arrangement.

However, although the prefractionator arrangement might require less energy than a conventional arrangement, all of the heat must be supplied at highest temperature and all of the heat rejected at the lowest temperature of separation. This can be particularly important if the distillation is at low temperature using refrigeration for condensation. From the comparison, the condenser temperature is much lower than that in the conventional arrangement; especially the depropaniser column condenser requires condensing temperature at 4°C. Therefore, a cost analysis should be carried out to ensure the new arrangement is more economical than the original arrangement based on the price given in Table 5.12.

Table 5.12: Price of utility

MP Steam	RM 28/ton
Cooling water	RM 0.13/m^3
Refrigeration @ 30°C	RM 62/kW/yr
Refrigeration @ 4°C	RM 115/kW/yr

Table 5.13: Cost comparison of conventional and prefractionator arrangements

		Conventional arrangement	Prefractionator arrangement
Depropaniser	Condenser	RM 840,000/yr	RM 141,330/yr
	Reboiler	RM 3,168,746/yr	RM 991,748/yr
	Annual cost	RM 4,008,748	RM 1,133,078
Debutaniser	Condenser	RM 484,335 /yr	RM 0/yr (*)
	Reboiler	RM 1,522,381/yr	RM 1,133,400/yr
	Annual cost	RM 2,006,717	RM 1,986,481
Annual total operation cost (energy)		RM 6,015,464	RM 2,266,481
Annual total savings (energy)		RM 0	**RM 3,748,983**

Note: (*) the operating cost of the condenser in the main column is taken care of by absorption chiller as it is at 4°C, which is within the range of operating temperature of absorption chiller systems. Therefore, this condenser is considered operating free.

Although the prefractionator arrangement uses refrigeration for both condensers, the overall savings from both reboilers and condensers exceed the cost of refrigeration utilization in the condensers. It could generate annual savings of around RM 3.7 million thanks to the new prefractionator arrangement. Most of the savings comes from 54% savings of MP steam usage in the new arrangement. Although the condensers generate 37% savings in term of energy, the penalty is on the utilization of refrigeration instead of cooling water as cooling medium for both condensers. Therefore, there is not much savings from condenser load.

Since the condensers of both prefractionator and depropaniser column operate at much lower temperature compared to the original arrangement; therefore, new pieces of heat exchanger are required to replace the existing ones. The estimated cost of new heat exchangers is calculated based on the heat transfer area and material used.

Table 5.14: Cost of new heat exchangers for condensers of Prefractionator and Depropaniser column due to the usage of refrigerant as cooling medium

Description	Surface area (m^2)	Material	Cost (RM)
Prefrac Condenser	345	KCS Shell; KCS Tube	1,402,010
De-C3 Condenser	4556	KCS Shell; KCS Tube	3,204,935

Some assumptions are made for the calculation of payback period as well as the rate of return for this investment as follow:

- Interest rate: 10%
- Assumed planned life of the project is 15 years
- NPV is calculated on the basis of 15 yrs
- Payback period is calculated based on HEs cost and piping modification only
- IRR is calculated based on estimated operating cost (15% of piping modification cost) and maintenance cost (5% of piping modification cost)

The detailed calculation for net present value (NPV) and internal rate of return (IRR) is attached in Appendix 10. The total investment is approximately RM 5.3 mil, which is the cost of two new heat exchangers and the piping modification cost (15% of the total cost of heat exchangers).

$$\text{Payback Period} = \frac{\text{Total investment cost}}{\text{Total savings}} = \frac{\text{RM 5,297,987}}{\text{RM 3,748,983}}$$

$$= 1.5 \text{ yrs}$$

Table 5.15: Summary of economic analysis for thermal coupling arrangement

NPV	RM 21.21mil
Payback period	1.5 yrs
IRR	35%

From economic evaluation, this improvement in PRU area has the great incentive for gas processors. Besides, there is opportunity for additional savings in this area by heat integration of product rundowns with upstream. The three products from the main column, which are propane, butane and condensates, need to be cooled down to required temperature. It is shown in Table 5.16.

Table 5.16: Utility load required for product rundowns

	Supply temp (°C)	Target Temp (°C)	Utility load (kW)
Propane	4.96	20.00	-340
Butane	49.17	21.00	341
Condensates	116.14	63.10	1089

Table 5.17: Source of heat integration

	Supply temp (°C)	Target Temp (°C)	Utility load (kW)
Compressed sale gas	67.00	69.60	340
Sale gas	17.00	19.38	-341
C2+ product	25.84	36.85	-1089

Propane product could be heat integrated with compressed sales gas to save both cooling water and refrigeration. Similarly, butane product can be heat integrated with sales gas and condensates product with preheated feed to de-ethaniser column (C2+ product). The savings is calculated in Table 5.18.

Table 5.18: Savings from heat integration of product rundowns

	Utility load (kW)	Predicted annual savings (RM)
Propane	340	21,002
Butane	341	41,871
Condensates	1089	133,716
Compressed sale gas	340	41,748
Sale gas	341	41,871
C2+ product	1089	67,268
Total savings		347,474

The heat integration of product rundowns with upstream could generate around RM350k savings per annum for the operating cost.

Chapter 6

CONCLUSIONS

Natural gas consumption is at high demand due to its outstanding characteristics. Therefore improvement of natural gas processing is a matter of great concern to many gas processors. This research has looked into the improvement of energy consumption in natural gas processing. The objectives of the research are to minimize the cold utility as well as the hot utility consumption and at the same time maximize the power generated from the turbo-expander. The scope of the research is focused on the low temperature separation (LTSU) and the product recovery (PRU) areas of the plant.

Pinch analysis has been carried out to find any opportunity to improve the energy consumption of the plant. For LTSU, attention has been given to improve power generated by increasing inlet temperature to turbo-expander, which results in 36% more power generated. Expansion valve is used to let down pressure of fluids without any heat recovery. Therefore, heaters are added after expansion valves in order to recover as much energy as possible after the let down pressure via expansion valves. By this method, the cold utility is reduced from 5040kW to 4041kW, generating 20% savings in energy consumption. However, the cold utility is required at lower refrigeration cooling level. In order to improve the cooling level, a constant cooling load of 2562kW at -12°C is introduced above the pinch as a cold stream. This will help to reduce the cooling load to 2172kW at -85.5°C. The improved cooling load as well as the additional power generated will generate predicted annual savings of RM1.2 million. Economic analysis has shown the incentive of this improvement that give payback period of 4.5yrs and internal rate of return (IRR) of 19%.

For PRU area, the composite curve for the current arrangement shows that there is no incentive to improve energy consumption due to the wide range of operating temperature in this area. The heat recovery in this area is very limited as shown by very little overlap between the hot and cold composite curves. Therefore, a new prefractionator arrangement is introduced to replace the conventional depropaniser and debutaniser arrangement. The new arrangement could save 54% in hot utility and 37% in cold utility, which is equivalent to RM 3.75million savings per annum for the gas plant. Economic evaluation has been carried out to prove efficiency of this new arrangement, with payback period of 1.5 years and internal rate of return of 35%

Besides that, there are some opportunities for heat integration between product rundowns and some loose end heaters and coolers, which could generate additional savings of around RM350k.

The objectives of the research have been achieved; the new gas plant arrangement could generate savings of up to RM4.95 million per annum. This is a great incentive for gas processors to improve their plant profitability.

Chapter 7

FUTURE WORK

This research can be further studied to design of a feasible gas plant to accommodate changes in gas plant margin such as change in product economic values such as propane, butane products. The plant should be able to respond to change in market demand of propane and butane products.

This research has been looking into the process side of gas processing plant alone, i.e. low temperature separation unit and product recovery unit. The utility has been assumed to be maintained as at original state in the Plant X. Therefore, for future work of this research, absorption chiller will be introduced into utility system together with refrigeration system to provide cooling required from the process side. These two systems will be integrated optimally to give the best performance to the plant requirement. Moreover, auto-refrigeration from turbo-expanders will also be maximized to reduce cold utility consumption in the cryogenic area.

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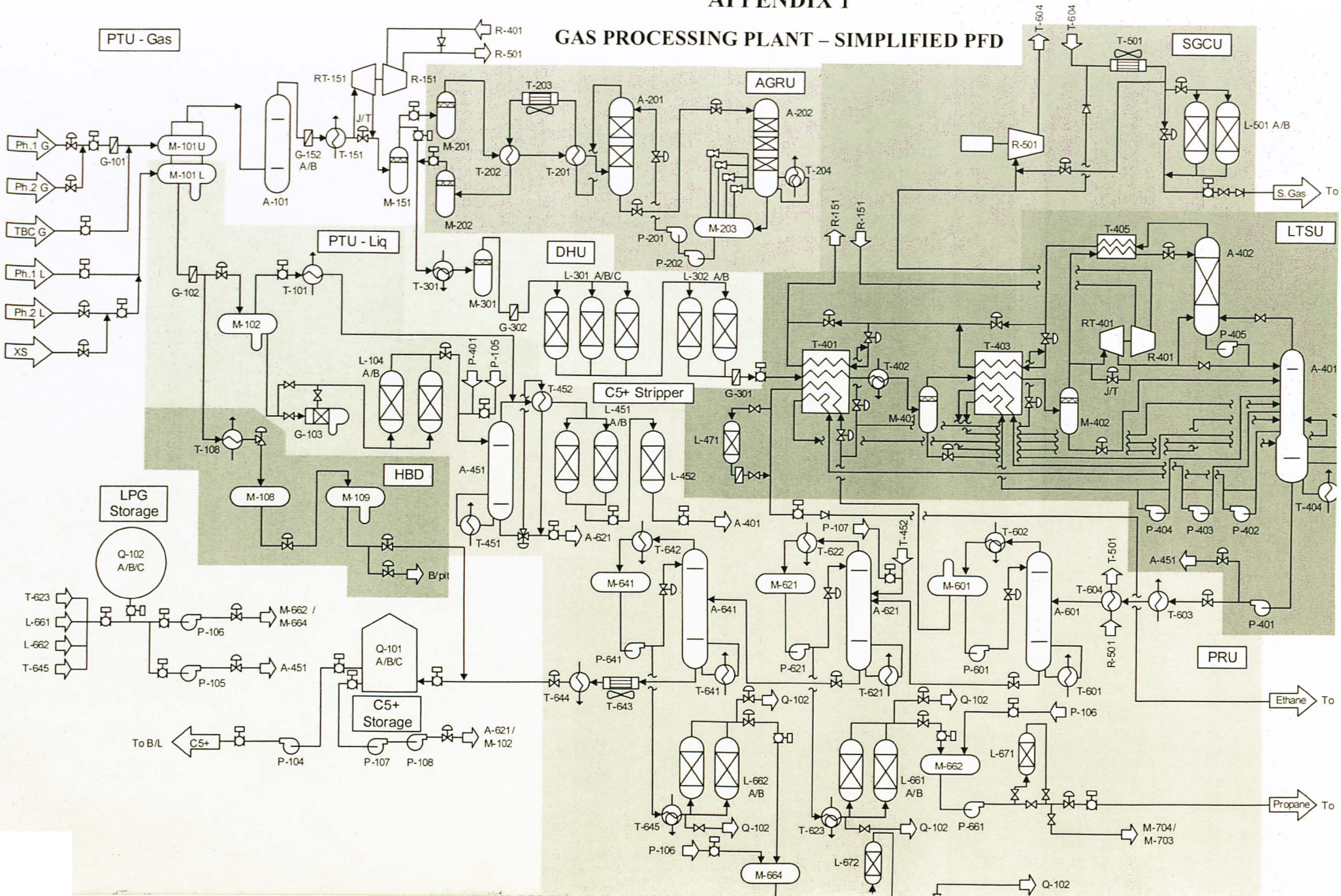
APPENDICES

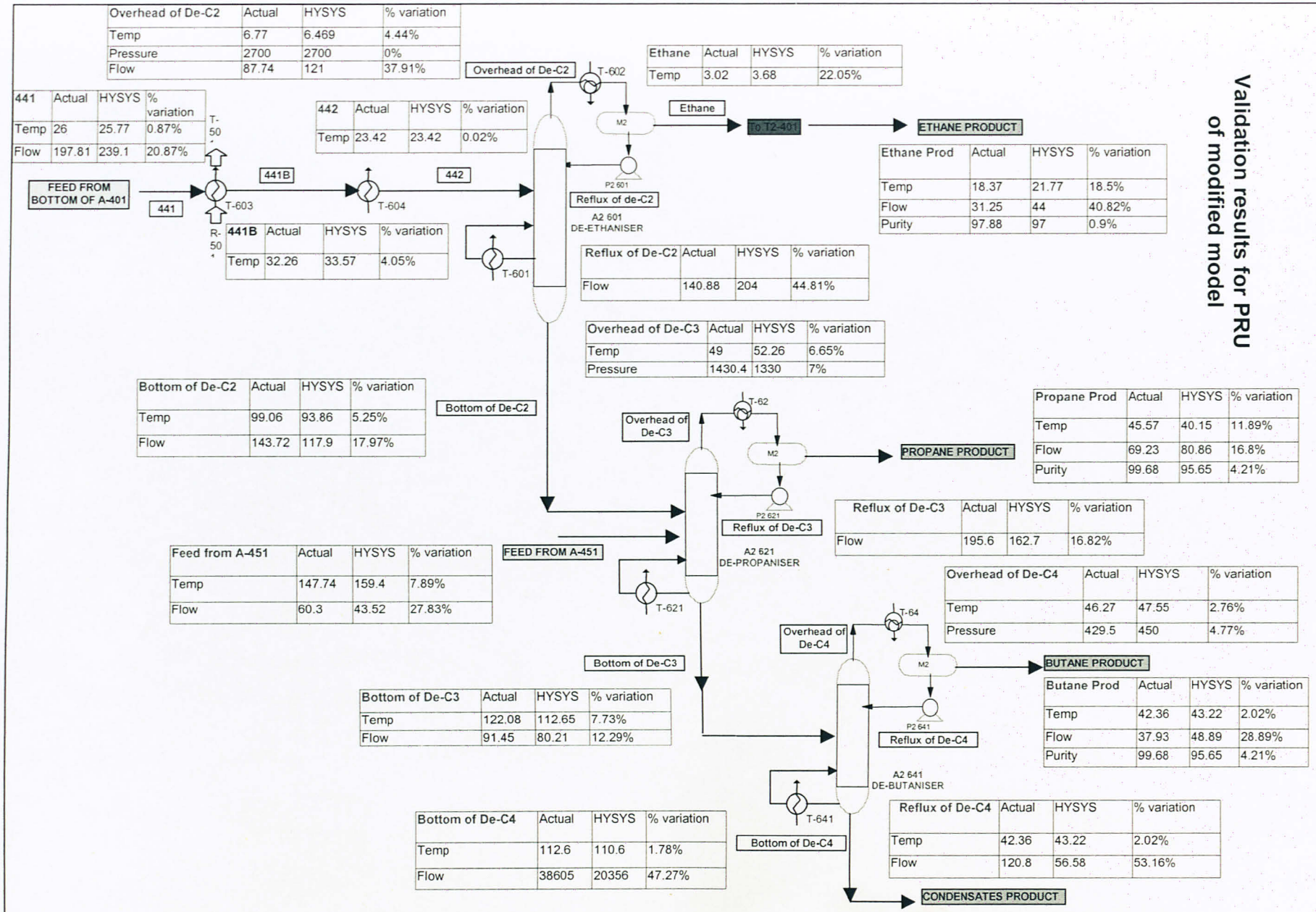
APPENDIX 1

GAS PROCESSING PLANT – SIMPLIFIED PFD



APPENDIX 2
DATA VALIDATION

Validation results for PRU of modified model



Overhead of De-C2	Actual	HYSYS	% variation
Temp	6.77	6.469	4.44%
Pressure	2700	2700	0%
Flow	87.74	121	37.91%

Ethane	Actual	HYSYS	% variation
Temp	3.02	3.68	22.05%

441	Actual	HYSYS	% variation
Temp	26	25.77	0.87%
Flow	197.81	239.1	20.87%

442	Actual	HYSYS	% variation
Temp	23.42	23.42	0.02%

Ethane Prod	Actual	HYSYS	% variation
Temp	18.37	21.77	18.5%
Flow	31.25	44	40.82%
Purity	97.88	97	0.9%

441B	Actual	HYSYS	% variation
Temp	32.26	33.57	4.05%

Reflux of De-C2	Actual	HYSYS	% variation
Flow	140.88	204	44.81%

Overhead of De-C3	Actual	HYSYS	% variation
Temp	49	52.26	6.65%
Pressure	1430.4	1330	7%

Bottom of De-C2	Actual	HYSYS	% variation
Temp	99.06	93.86	5.25%
Flow	143.72	117.9	17.97%

Propane Prod	Actual	HYSYS	% variation
Temp	45.57	40.15	11.89%
Flow	69.23	80.86	16.8%
Purity	99.68	95.65	4.21%

Reflux of De-C3	Actual	HYSYS	% variation
Flow	195.6	162.7	16.82%

Feed from A-451	Actual	HYSYS	% variation
Temp	147.74	159.4	7.89%
Flow	60.3	43.52	27.83%

Overhead of De-C4	Actual	HYSYS	% variation
Temp	46.27	47.55	2.76%
Pressure	429.5	450	4.77%

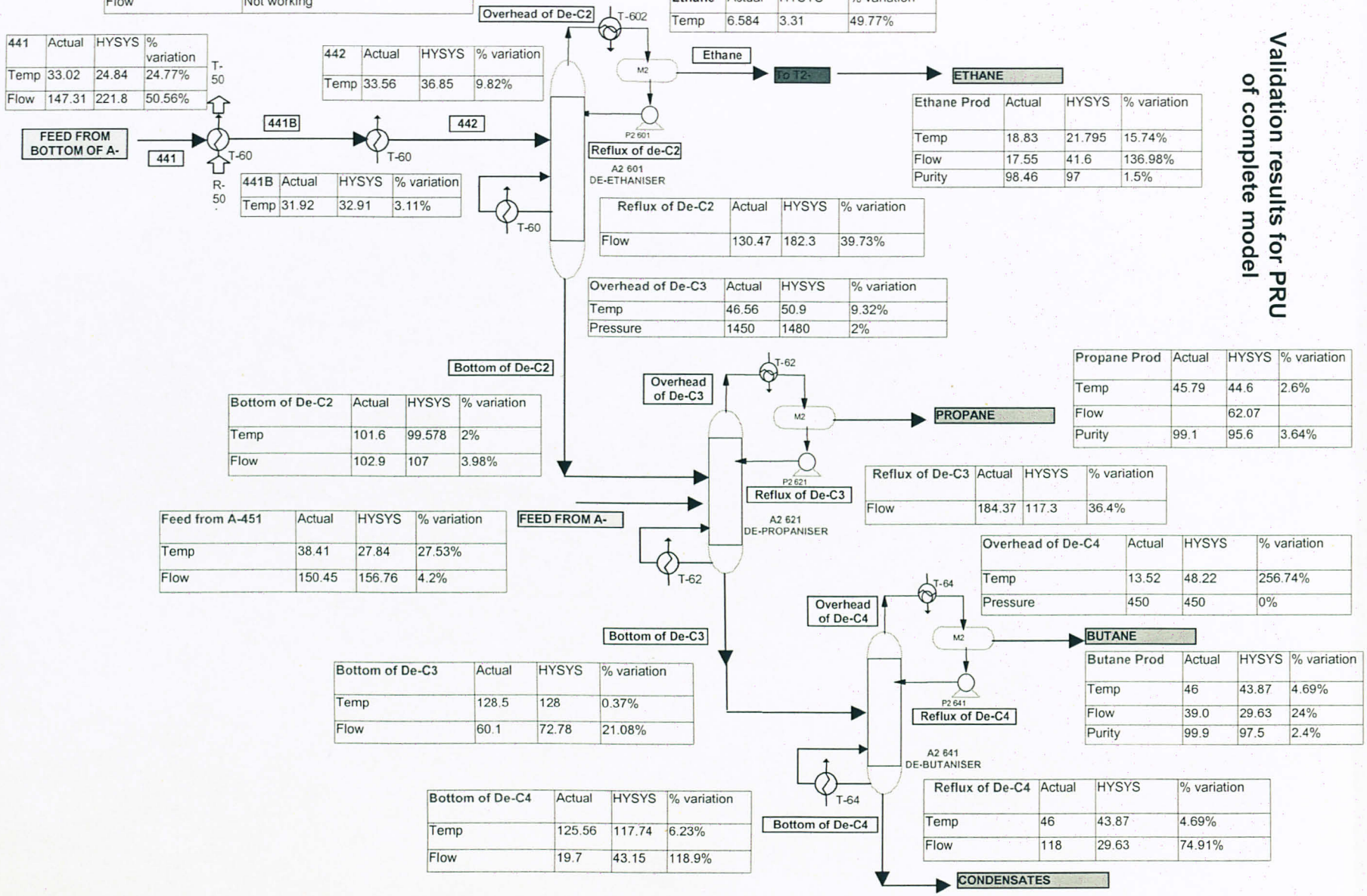
Bottom of De-C3	Actual	HYSYS	% variation
Temp	122.08	112.65	7.73%
Flow	91.45	80.21	12.29%

Butane Prod	Actual	HYSYS	% variation
Temp	42.36	43.22	2.02%
Flow	37.93	48.89	28.89%
Purity	99.68	95.65	4.21%

Bottom of De-C4	Actual	HYSYS	% variation
Temp	112.6	110.6	1.78%
Flow	38605	20356	47.27%

Reflux of De-C4	Actual	HYSYS	% variation
Temp	42.36	43.22	2.02%
Flow	120.8	56.58	53.16%

Validation results for PRU of complete model



Overhead of De-C2	Actual	HYSYS	% variation
Temp	5.683	6.469	13.82%
Pressure	2700	2700	0
Flow	Not working		

Ethane	Actual	HYSYS	% variation
Temp	6.584	3.31	49.77%

441	Actual	HYSYS	% variation
Temp	33.02	24.84	24.77%
Flow	147.31	221.8	50.56%

442	Actual	HYSYS	% variation
Temp	33.56	36.85	9.82%

Ethane Prod	Actual	HYSYS	% variation
Temp	18.83	21.795	15.74%
Flow	17.55	41.6	136.98%
Purity	98.46	97	1.5%

441B	Actual	HYSYS	% variation
Temp	31.92	32.91	3.11%

Reflux of De-C2	Actual	HYSYS	% variation
Flow	130.47	182.3	39.73%

Overhead of De-C3	Actual	HYSYS	% variation
Temp	46.56	50.9	9.32%
Pressure	1450	1480	2%

Propane Prod	Actual	HYSYS	% variation
Temp	45.79	44.6	2.6%
Flow		62.07	
Purity	99.1	95.6	3.64%

Bottom of De-C2	Actual	HYSYS	% variation
Temp	101.6	99.578	2%
Flow	102.9	107	3.98%

Reflux of De-C3	Actual	HYSYS	% variation
Flow	184.37	117.3	36.4%

Feed from A-451	Actual	HYSYS	% variation
Temp	38.41	27.84	27.53%
Flow	150.45	156.76	4.2%

Overhead of De-C4	Actual	HYSYS	% variation
Temp	13.52	48.22	256.74%
Pressure	450	450	0%

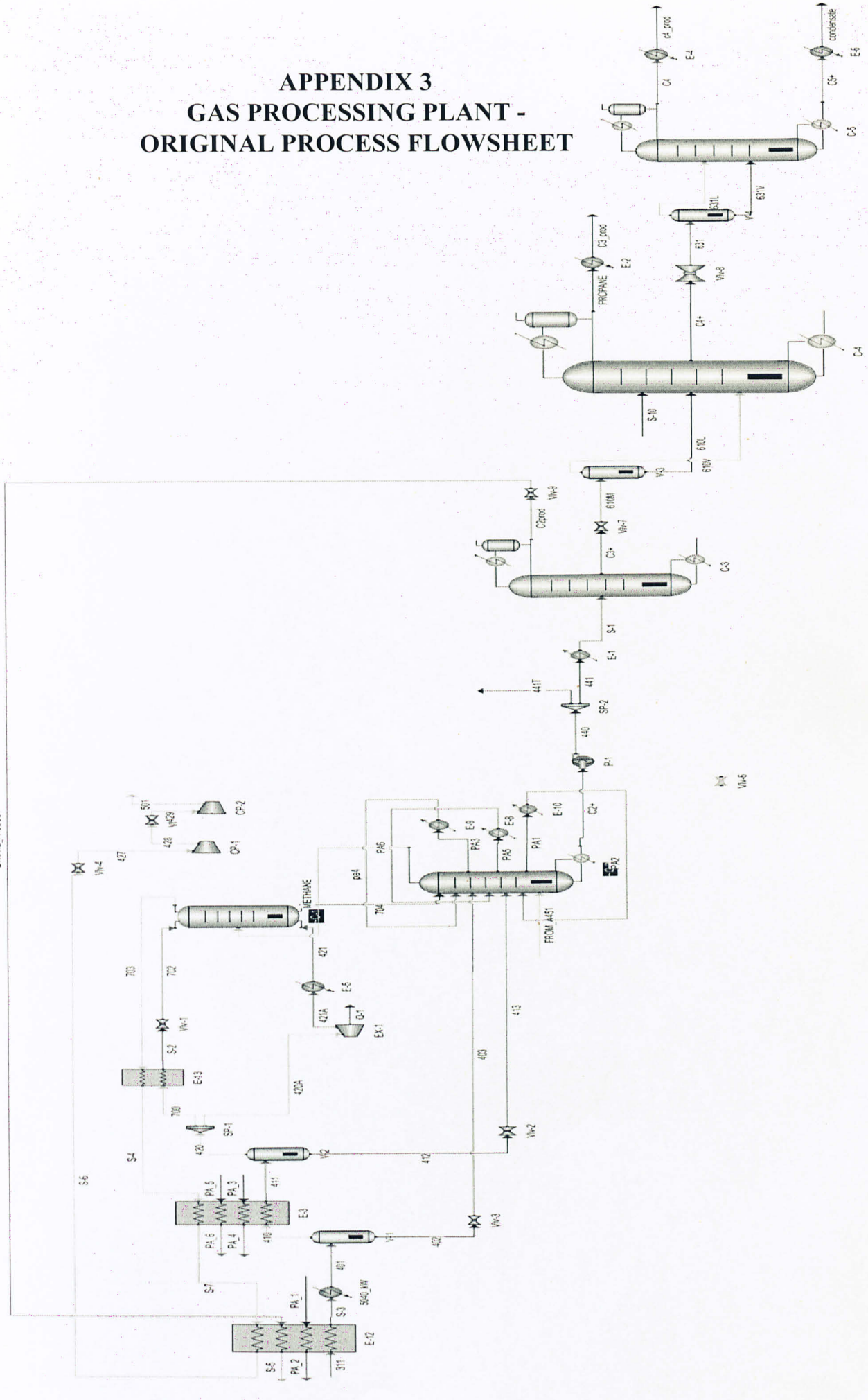
Bottom of De-C3	Actual	HYSYS	% variation
Temp	128.5	128	0.37%
Flow	60.1	72.78	21.08%

Butane Prod	Actual	HYSYS	% variation
Temp	46	43.87	4.69%
Flow	39.0	29.63	24%
Purity	99.9	97.5	2.4%

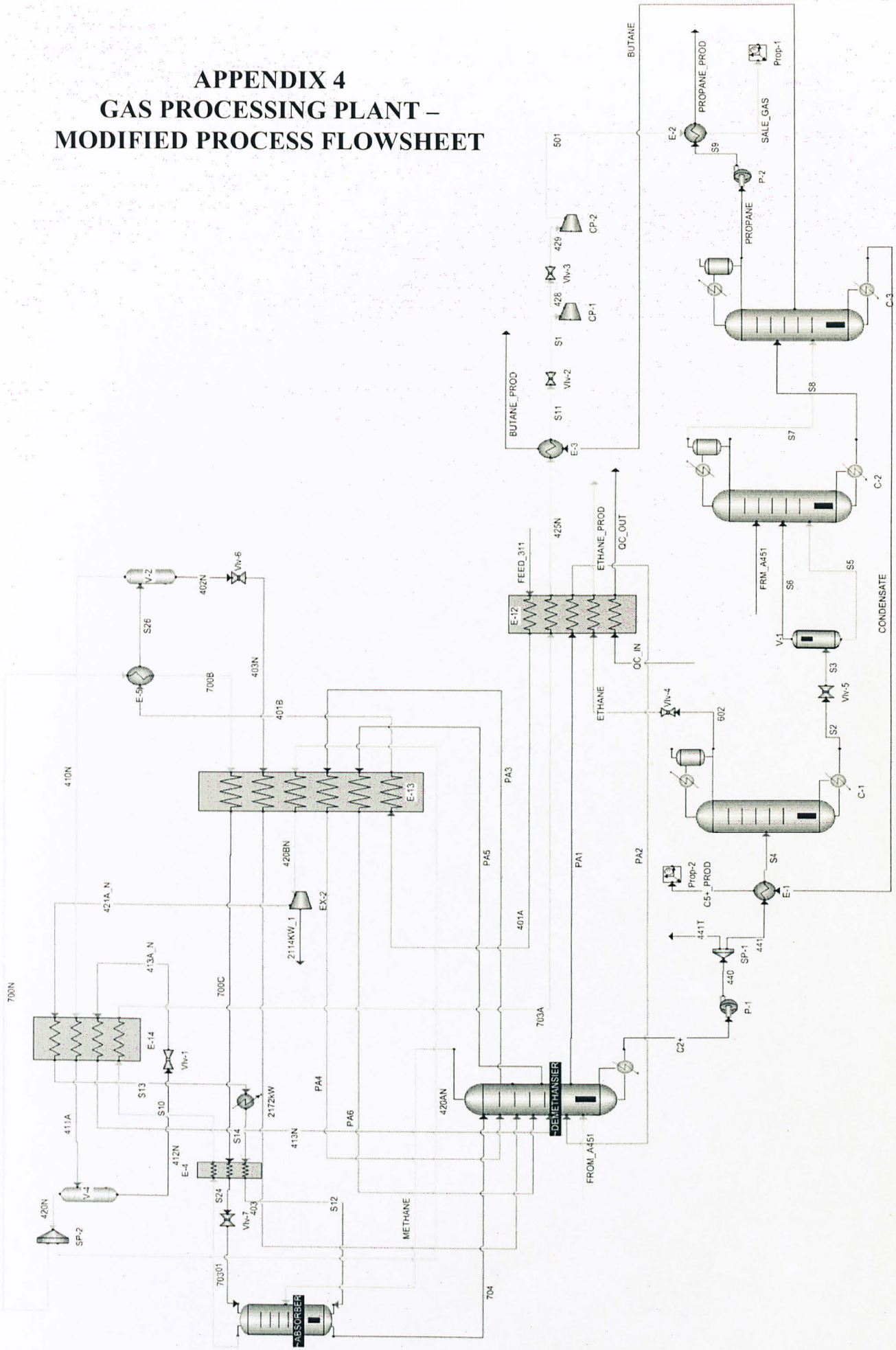
Bottom of De-C4	Actual	HYSYS	% variation
Temp	125.56	117.74	6.23%
Flow	19.7	43.15	118.9%

Reflux of De-C4	Actual	HYSYS	% variation
Temp	46	43.87	4.69%
Flow	118	29.63	74.91%

APPENDIX 3
GAS PROCESSING PLANT - ORIGINAL PROCESS FLOWSHEET



APPENDIX 4
GAS PROCESSING PLANT – MODIFIED PROCESS FLOWSHEET



Absorber: /ABSORBER

Ports

	Feed_0_overhea	Feed_6_bottc	Feed_6_feed	LiquidDraw_6_bo	overheadV
Connected To	/701.Out	/S12.Out	/METHANE.Ou	/704.In	/703.In
VapFrac	0	0.82940006	1	0	1
T [C]	-100.5858685	-89	-94.13076602	-94.29957548	-101.3703492
P [kPa]	2426	2370	2368	2368	2250
MoleFlow [kgmole/h]	4411.986853	8193.68987	5264.505411	4646.36172	13223.82041
MassFlow [kg/h]	77577.1062	144071.769	86024.43238	92876.95146	214796.3558
Energy [kJ/h]	-6577687.579	28667844.2	24812338.39	-12043957.51	58946452.48
Cp [kJ/kmol-K]	81.20761202	52.7669798	51.47736307	73.66982801	53.31805765
Mole Fraction					
METHANE	0.909281854	0.90928185	0.980402823	0.774232247	0.985047103
ETHANE	0.063525799	0.0635258	0.015456555	0.175273545	0.005125033
PROPANE	0.013929307	0.01392931	0.00051149	0.037860193	0.000179141
n-BUTANE	0.000840492	0.00084049	4.54E-06	0.00228096	1.57E-06
ISOBUTANE	0.001132898	0.0011329	1.09E-05	0.003074546	4.01E-06
ISOPENTANE	0.000155096	0.0001551	1.58E-07	0.000420813	5.05E-08
n-PENTANE	6.08E-05	6.08E-05	5.62E-08	0.000164962	1.96E-08
n-HEXANE	3.15E-06	3.15E-06	4.65E-10	8.54E-06	1.93E-10
n-HEPTANE	2.86E-07	2.86E-07	7.13E-12	7.75E-07	2.68E-12
n-OCTANE	7.14E-09	7.14E-09	3.48E-14	1.94E-08	1.24E-14
n-NONANE	2.03E-10	2.03E-10	2.03E-16	5.52E-10	7.76E-17
n-DECANE	7.04E-12	7.04E-12	1.07E-18	1.91E-11	3.93E-19
n-UNDECANE	0	0	1.00E-100	1.00E-100	1.00E-100
n-DODECANE	0	0	1.00E-100	1.00E-100	1.00E-100
n-TRIDECANE	0	0	1.00E-100	1.00E-100	1.00E-100
n-TETRADECANE	0	0	1.00E-100	1.00E-100	1.00E-100
CARBON DIOXIDE	0.001942143	0.00194214	0.001433335	0.005453507	0.000505819
NITROGEN	0.00912817	0.00912817	0.002180123	0.001229894	0.009137261

Tower	Stage	Feed/Draw	EnergyFeed	T	P
		kgmole/h	kJ/h	C	kPa
VF	1	V=13223.82 F=4411.99		-101.3703492	2250
	2			-101.0083392	2269.666667
	3			-100.6794946	2289.333333
	4			-100.2478378	2309
	5			-99.51615304	2328.666667
	6			-97.97794703	2348.333333
LFF	7	L=4646.36 F=5264.51 F=8193.(		-94.29957548	2368

		In	Out
Connected To		/S1.Out	/428.In
VapFrac		1	1
T [C]		18.83777647	29.6983479
P [kPa]		2091.5	2358
MoleFlow [kgmole/h]		13223.82041	13223.8204
MassFlow [kg/h]		214796.3558	214796.356
Energy [kJ/h]		124581129.5	129530632
Cp [kJ/kmol-K]		38.74807157	39.2207254
Mole Fraction			
METHANE		0.985047103	0.9850471
ETHANE		0.005125033	0.00512503
PROPANE		0.000179141	0.00017914
n-BUTANE		1.57E-06	1.57E-06
ISOBUTANE		4.01E-06	4.01E-06
ISOPENTANE		5.05E-08	5.05E-08
n-PENTANE		1.96E-08	1.96E-08
n-HEXANE		1.93E-10	1.93E-10
n-HEPTANE		2.68E-12	2.68E-12
n-OCTANE		1.24E-14	1.24E-14
n-NONANE		7.76E-17	7.76E-17
n-DECANE		3.93E-19	3.93E-19
n-UNDECANE		1.00E-100	1.00E-100
n-DODECANE		1.00E-100	1.00E-100
n-TRIDECANE		1.00E-100	1.00E-100
n-TETRADECANE		1.00E-100	1.00E-100
CARBON DIOXIDE		0.000505819	0.00050582
NITROGEN		0.009137261	0.00913726

InQ		Work [W]	1374861.729
DeltaP		DP [Kpa]	266.5
PolytropicHead		Head [m]	1770.550796
AdiabaticExponent		Generic	1.302577376
PolytropicExponent		Generic	1.449151511
AdiabaticEff		Generic	0.75
PolytropicEff		Generic	0.753519424
AdiabaticHead		Head [m]	1762.281177
PolytropicHeadFactor		Generic	0.997968059

ValveDP: /Vlv-1

	In	Out
Connected To	/412N.Out	/413A_N.In
VapFrac	0	0.330194279
T [C]	-50.0000005	-70.78387117
P [kPa]	5380	2575
MoleFlow [kgmol	590.568757	590.5687569
MassFlow [kg/h]	15138.5066	15138.50657
Energy [kJ/h]	-1.05E+06	-1.05E+06
Cp [kJ/kmol-K]	8.58E+01	6.71E+01

DP [Kpa] 2805

VapFrac	1	1
T [C]	19.3843915	18.83777647
P [kPa]	2203	2091.5
MoleFlow [kgmol	13223.8204	13223.82041
MassFlow [kg/h]	214796.356	214796.3558
Energy [kJ/h]	124581129	124581129.5
Cp [kJ/kmol-K]	38.8903217	38.74807157

DP [Kpa] 111.5

ValveDP: /Vlv-3

	In	Out
Connected To	/428.Out	/429.In
VapFrac	1	1
T [C]	29.6983479	29.29298068
P [kPa]	2358	2269
MoleFlow [kgmol	13223.8204	13223.82041
MassFlow [kg/h]	214796.356	214796.3558
Energy [kJ/h]	129530632	129530631.7
Cp [kJ/kmol-K]	39.2207254	39.11632655

DP [Kpa] 89

ValveDP: /Vlv-4

	In	Out
Connected To	/602.Out	/ETHANE.In
VapFrac	0	0.288664453
T [C]	3.14690827	-27.83030011
P [kPa]	2770	1200
MoleFlow [kgmol	1376.74646	1376.746465
MassFlow [kg/h]	41752.2257	41752.22571
Energy [kJ/h]	-407655.597	-407655.5966
Cp [kJ/kmol-K]	134.065707	82.49116838

DP [Kpa] 1570

ValveDP: /Vlv-5

	In	Out
Connected To	/S2.Out	/S3.In
VapFrac	1.71E-08	0.361347831
T [C]	100.155888	73.66648457
P [kPa]	2835	1600
MoleFlow [kgmol	1106.75922	1106.759217
MassFlow [kg/h]	58745.8048	58745.80479
Energy [kJ/h]	10288932.6	10288932.59
Cp [kJ/kmol-K]	209.888578	145.9849442

DP [Kpa] 1235

ValveDP: /Vlv-6

	In	Out

VapFrac	1	1
T [C]	19.3843915	18.83777647
P [kPa]	2203	2091.5
MoleFlow [kgmol	13223.8204	13223.82041
MassFlow [kg/h]	214796.356	214796.3558
Energy [kJ/h]	124581129	124581129.5
Cp [kJ/kmol-K]	38.8903217	38.74807157

DP [Kpa] 111.5

ValveDP: /Vlv-3

	In	Out
Connected To	/428.Out	/429.In
VapFrac	1	1
T [C]	29.6983479	29.29298068
P [kPa]	2358	2269
MoleFlow [kgmol	13223.8204	13223.82041
MassFlow [kg/h]	214796.356	214796.3558
Energy [kJ/h]	129530632	129530631.7
Cp [kJ/kmol-K]	39.2207254	39.11632655

DP [Kpa] 89

ValveDP: /Vlv-4

	In	Out
Connected To	/602.Out	/ETHANE.In
VapFrac	0	0.288664453
T [C]	3.14690827	-27.83030011
P [kPa]	2770	1200
MoleFlow [kgmol	1376.74646	1376.746465
MassFlow [kg/h]	41752.2257	41752.22571
Energy [kJ/h]	-407655.597	-407655.5966
Cp [kJ/kmol-K]	134.065707	82.49116838

DP [Kpa] 1570

ValveDP: /Vlv-5

	In	Out
Connected To	/S2.Out	/S3.In
VapFrac	1.71E-08	0.361347831
T [C]	100.155888	73.66648457
P [kPa]	2835	1600
MoleFlow [kgmol	1106.75922	1106.759217
MassFlow [kg/h]	58745.8048	58745.80479
Energy [kJ/h]	10288932.6	10288932.59
Cp [kJ/kmol-K]	209.888578	145.9849442

DP [Kpa] 1235

T [C]	-36.5000002	-53.9337108
P [kPa]	5445	2468
MoleFlow [kgmol	2493.75452	2493.75452
MassFlow [kg/h]	79923.7836	79923.78357
Energy [kJ/h]	-7383497.2	-7383497.197
Cp [kJ/kmol-K]	89.2595718	75.35649443

DP [Kpa] 2977

ValveDP: /Vlv-7

	In	Out
Connected To	/S24.Out	/701.In
VapFrac	0	0
T [C]	-99.5	-100.5858685
P [kPa]	5360	2426
MoleFlow [kgmol	4411.98685	4411.986853
MassFlow [kg/h]	77577.1062	77577.1062
Energy [kJ/h]	-6577687.58	-6577687.579
Cp [kJ/kmol-K]	73.1346155	81.20761202

DP [Kpa] 2934

ExpanderWithCurve: /EX-2

Ports

		In	Out
Connected To		/420BN.Out	/421A_N.In
VapFrac		1	0.97424065
T [C]		-30	-68.0113004
P [kPa]		5370	2375
MoleFlow [kgmole/h]		8193.68987	8193.68987
MassFlow [kg/h]		144071.7686	144071.769
Energy [kJ/h]		52228348.63	45442101.7
Cp [kJ/kmol-K]		55.35654874	45.3843278

Ports	ConnectedTo		
OutQ	/2114KW_1.In	Work [W]	1885068.592
DeltaP		DP [Kpa]	2995
PolytropicHead		Head [m]	6533.655033
AdiabaticExponent		Generic	1.197910501
ExpanderSpeed			
PolytropicHeadFactor		Generic	0.985272763
AdiabaticEff		Generic	0.75
PolytropicEff		Generic	0.735146501
AdiabaticHead		Head [m]	6404.258179
PolytropicExponent		Generic	1.130227085

VapFrac		0.943505278	0.87459222
T [C]		-76	-85.5
P [kPa]		2370	2370
MoleFlow [kgmole/h]		8193.68987	8193.68987
MassFlow [kg/h]		144071.7686	144071.769
Energy [kJ/h]		40258358	32438284.2
Cp [kJ/kmol-K]		47.25453201	50.7688328
Mole Fraction			
METHANE		0.909281854	0.90928185
ETHANE		0.063525799	0.0635258
PROPANE		0.013929307	0.01392931
n-BUTANE		0.000840492	0.00084049
ISOBUTANE		0.001132898	0.0011329
ISOPENTANE		0.000155096	0.0001551
n-PENTANE		6.08E-05	6.08E-05
n-HEXANE		3.15E-06	3.15E-06
n-HEPTANE		2.86E-07	2.86E-07
n-OCTANE		7.14E-09	7.14E-09
n-NONANE		2.03E-10	2.03E-10
n-DECANE		7.04E-12	7.04E-12
n-UNDECANE		0	0
n-DODECANE		0	0
n-TRIDECANE		0	0
n-TETRADECANE		0	0
CARBON DIOXIDE		0.001942143	0.00194214
NITROGEN		0.00912817	0.00912817

OutQ	Energy [kJ/h]	7820073.8
DeltaP	DP [Kpa]	0

Segment	T [C]	EnergyAcun	H [kJ/kmol]
0	-76.00000021	0	4913.33679
1	-85.49999994	7820073.8	3958.934822

PumpWithCurve: /P-1

Ports		In	Out
Connected To		/C2+.Out	/440.In
VapFrac		3.86E-10	0
T [C]		25	25.8372967
P [kPa]		2396	3009
MoleFlow [kgmole/h]		2.66E+03	2.66E+03
MassFlow [kg/h]		107579.615	107579.615
Energy [kJ/h]		-1010734.66	-825931.747
Cp [kJ/kmol-K]		125.7750389	124.275424

Ports	ConnectedTo		
InQ		Work [W]	51334.14231
DeltaP		DP [Kpa]	613
Efficiency		Generic	0.75
Head		Length [m]	133.9938048

P [kPa]		550	1450
MoleFlow [kgmole/h]		695.8335236	695.833524
MassFlow [kg/h]		30689.80768	30689.8077
Energy [kJ/h]		-2590140.421	-2523583.63
Cp [kJ/kmol-K]		113.7470913	113.085055

CompressorWithCurve: /CP-2

		In	Out
Connected To		/429.Out	/501.In
VapFrac		1	1
T [C]		29.29298068	69.560607
P [kPa]		2269	3467
MoleFlow [kgmole/h]		13223.82041	13223.8204
MassFlow [kg/h]		214796.3558	214796.356
Energy [kJ/h]		129530631.7	148362162
Cp [kJ/kmol-K]		39.11632655	40.9956477
Mole Fraction			
METHANE		0.985047103	0.9850471
ETHANE		0.005125033	0.00512503
PROPANE		0.000179141	0.00017914
n-BUTANE		1.57E-06	1.57E-06
ISOBUTANE		4.01E-06	4.01E-06
ISOPENTANE		5.05E-08	5.05E-08
n-PENTANE		1.96E-08	1.96E-08
n-HEXANE		1.93E-10	1.93E-10
n-HEPTANE		2.68E-12	2.68E-12
n-OCTANE		1.24E-14	1.24E-14
n-NONANE		7.76E-17	7.76E-17
n-DECANE		3.93E-19	3.93E-19
n-UNDECANE		1.00E-100	1.00E-100
n-DODECANE		1.00E-100	1.00E-100
n-TRIDECANE		1.00E-100	1.00E-100
n-TETRADECANE		1.00E-100	1.00E-100
CARBON DIOXIDE		0.000505819	0.00050582
NITROGEN		0.009137261	0.00913726

DistillationColumn: /C-1

Head		Length [m]	173.7189136

InQ		Work [W]	5230980.713
DeltaP		DP [Kpa]	1198
PolytropicHead		Head [m]	6809.853527
AdiabaticExponent		Generic	1.301163646
CompressorSpeed			
PolytropicExponent		Generic	1.435331875
AdiabaticEff		Generic	0.75
PolytropicEff		Generic	0.761727677
AdiabaticHead		Head [m]	6705.007971
PolytropicHeadFactor		Generic	0.998647668

T [C]		36.85	3.14690827	100.1558882	3.146908267
P [kPa]		2936	2770	2835	2770
MoleFlow [kgmole/h]		2483.505681	1376.74646	1106.759217	1.00E-40
MassFlow [kg/h]		100498.0305	41752.2257	58745.80479	3.04E-39
Energy [kJ/h]		3149090.564	-407655.597	10288932.59	0
Cp [kJ/kmol-K]		135.9894061	134.065707	209.8885775	82.17639943
Mole Fraction					
METHANE		0.003514872	0.00634046	1.73E-13	0.021111675
ETHANE		0.538860076	0.9688	0.004038895	0.932952039
PROPANE		0.257624861	0.00089288	0.576985068	0.000337658
n-BUTANE		0.058178454	1.44E-10	0.130549191	2.29E-11
ISOBUTANE		0.052333218	1.39E-09	0.117432808	2.52E-10
ISOPENTANE		0.04029097	1.07E-15	0.090410679	8.23E-17
n-PENTANE		0.019667141	4.19E-17	0.044131963	2.69E-18
n-HEXANE		0.00959136	4.71E-25	0.021522474	1.02E-26
n-HEPTANE		0.005525656	3.09E-30	0.012399263	3.38E-32
n-OCTANE		0.000900423	5.20E-36	0.002020498	2.87E-38
n-NONANE		0.000177519	9.59E-43	0.000398342	2.21E-45
n-DECANE		4.89E-05	3.16E-49	0.00010968	3.19E-52
n-UNDECANE		3.62E-07	6.58E-57	8.12E-07	2.93E-60
n-DODECANE		1.45E-07	1.17E-62	3.25E-07	2.42E-66
n-TRIDECANE		1.00E-100	1.00E-100	1.00E-100	1.00E-100
n-TETRADECANE		1.00E-100	1.00E-100	1.00E-100	1.00E-100
CARBON DIOXIDE		0.013286066	0.02396666	3.54E-09	0.045598614
NITROGEN		9.97E-10	1.80E-09	1.70E-25	1.32E-08

	5			6.039252816	2777.222222
	6			6.424416687	2779.027778
	7			6.971702623	2780.833333
	8			7.751655002	2782.638889
	9			8.838121096	2784.444444
	10			10.29267782	2786.25
	11			12.13561206	2788.055556
	12			14.3164238	2789.861111
	13			16.7104249	2791.666667
	14			19.16466585	2793.472222
	15			21.59041745	2795.277778
	16			24.09340081	2797.083333
	17			27.22412318	2798.888889
F	18	F=2483.51		32.9019939	2800.694444
	19			33.81298364	2802.5
	20			34.82691625	2804.305556
	21			36.22411611	2806.111111
	22			38.27255616	2807.916667
	23			41.23056674	2809.722222
	24			45.26755761	2811.527778
	25			50.3307573	2813.333333
	26			56.05925331	2815.138889
	27			61.86942464	2816.944444
	28			67.19228897	2818.75
	29			71.67969553	2820.555556
	30			75.25000775	2822.361111
	31			78.01495644	2824.166667
	32			80.191814	2825.972222
	33			82.06227802	2827.777778
	34			83.99947539	2829.583333
	35			86.59537505	2831.388889
	36			91.01988685	2833.194444
LQ*	37 (reboiler	L*=1106.76	Q=3.6781e+0	100.1558882	2835

Energy [kJ/h]	3122599.027	7166333.55	5264454.471	-0.256679808	5267728.507	11774764
Cp [kJ/kmol-K]	167.3318595	108.256051	264.2752047	132.2031257	215.8456706	86.716947
Mole Fraction						
METHANE	1.73E-13	1.73E-13	0	1.39E-14	5.90E-43	2.36E-13
ETHANE	0.002229468	0.00723691	0.010770108	0.002222287	1.11E-10	0.0082865
PROPANE	0.493161731	0.72513583	0.215482155	0.686302187	0.000916795	0.8420561
n-BUTANE	0.148743743	0.09839184	0.073460735	0.147657213	0.216811883	0.0621788
ISOBUTANE	0.126963469	0.10058816	0.08601086	0.162818312	0.152880667	0.0872843
ISOPENTANE	0.117098064	0.04324295	0.078070781	0.000867328	0.230388599	0.0001706
n-PENTANE	0.058057138	0.01952038	0.04304043	0.000132672	0.114599378	2.38E-05
n-HEXANE	0.031196046	0.00442524	0.076610766	6.61E-11	0.078965947	3.34E-12
n-HEPTANE	0.018658702	0.00133623	0.105071051	9.50E-16	0.070755073	1.90E-17
n-OCTANE	0.003101487	0.00010994	0.112861129	2.41E-21	0.051221038	1.83E-23
n-NONANE	0.000617482	1.10E-05	0.081510815	9.81E-27	0.034666098	2.92E-29
n-DECANE	0.000170894	1.49E-06	0.048420484	3.25E-32	0.020314527	3.45E-35
n-UNDECANE	1.27E-06	5.67E-09	0.023830238	5.99E-39	0.009881223	2.62E-42
n-DODECANE	5.08E-07	1.18E-09	0.022170222	8.47E-44	0.009191952	1.52E-47
n-TRIDECANE	1.00E-100	1.00E-100	0.011630116	1.35E-91	0.004821566	9.54E-96
n-TETRADECANE	1.00E-100	1.00E-100	0.011060111	1.00E-100	0.004585255	1.00E-100
CARBON DIOXIDE	1.28E-09	7.53E-09	0	5.97E-10	9.74E-32	4.82E-09
NITROGEN	1.70E-25	1.70E-25	0	5.07E-27	2.64E-65	2.32E-25

	Ports	ConnectedTo	
Ene In	EnergyFeed_39_reboilerQ	Energy [kJ/h]	9725395.008
Ene Out	EnergyFeed_0_condenserQ	Energy [kJ/h]	8236289.63
Signal	Variable_0_Reflux	Generic	0.598338458
	Variable_39_MoleRatio	Fraction	0.008
	Variable_0_MoleFraction	Fraction	0.001
	Variable_0_Stage_T	T [C]	30.74081323
	Variable_39_MoleRatio_1	Fraction	0.003979342
	DegSubCool_0_dsc	DT [C]	0

Tower	Stage	Feed/Draw	EnergyFeed	T	P	Liquid Flow	Vapour Flow
		kgmole/h	kJ/h	C	kPa	kgmole/h	kgmole/h
LVQ**	1 (condens	L**=0.00 V=811.3	Q=-8.2363e+	30.74081323	850	485.4602135	811.34696
	2			34.52678647	851.2820513	474.3305339	1296.8074
	3			36.57058615	852.5641026	468.723363	1285.6777
	4			37.66704939	853.8461538	465.2744507	1280.0705
	5			38.32548455	855.1282051	462.4157931	1276.6216

	8			39.91645082	858.974359	449.9700996	1266.6651
	9			40.68917075	860.2564103	442.8185549	1261.3173
	10			41.70701827	861.5384615	433.4691079	1254.1657
	11			43.03357008	862.8205128	420.9254502	1244.8163
	12			44.81833374	864.1025641	399.1391598	1232.2726
FF	13	F=706.83 F=399.93		47.93092188	865.3846154	894.3034381	1210.4863
	14			48.41425256	866.6666667	894.3207289	598.89139
	15			49.06267546	867.9487179	894.2195512	598.90868
	16			49.93331293	869.2307692	894.0731152	598.8075
	17			51.0800185	870.5128205	894.0065995	598.66107
	18			52.54124464	871.7948718	894.2226846	598.59455
	19			54.3215408	873.0769231	894.9808508	598.81064
	20			56.37533252	874.3589744	896.512195	599.5688
	21			58.6061074	875.6410256	898.8845817	601.10015
	22			60.89144076	876.9230769	901.8679145	603.47253
	23			63.13490505	878.2051282	904.7819738	606.45587
	24			65.35191005	879.4871795	905.6070259	609.36992
	25			67.96036029	880.7692308	883.2231222	610.19498
F	26	F=209.20		76.50377326	882.0512821	912.8817774	587.81107
	27			78.13620781	883.3333333	920.3896427	408.26973
	28			79.65116619	884.6153846	927.4640743	415.77759
	29			81.03246909	885.8974359	934.1111977	422.85202
	30			82.25409651	887.1794872	940.1791892	429.49915
	31			83.30296687	888.4615385	945.5302678	435.56714
	32			84.18434105	889.7435897	950.1012965	440.91822
	33			84.91980689	891.025641	953.9027439	445.48925
	34			85.54479583	892.3076923	956.9867	449.29069
	35			86.11141333	893.5897436	959.3960477	452.37465
	36			86.70466274	894.8717949	961.076589	454.784
	37			87.49133236	896.1538462	961.6305207	456.46454
	38			88.86628198	897.4358974	959.0041777	457.01847
	39			92.04962083	898.7179487	936.9269238	454.39213
LQ*	40 (reboiler	L*=504.61	Q=9.7254e+0	103.7530791	900	504.6120494	432.31487

DistillationColumn: /C-3

Ports

		Feed_11_feed	Feed_28_fee	condenserL	LiquidDraw_19_L	reboilerL
Connected To		/S7.Out	/S8.Out	/PROPANE.In	/BUTANE.In	/CONDENSATE.In
VapFrac		1	4.54E-09	1.71E-09	0	2.38E-08
T [C]		30.74081323	103.753079	4.181559845	49.17332427	116.1432183
P [kPa]		850	900	550	575	600
MoleFlow [kgmole/h]		811.3469612	504.612049	695.8335236	295.9875663	324.1379207
MassFlow [kg/h]		37287.92089	39566.1948	30629.80768	17237.51785	29026.7002

Energy [kJ/h]	11774764.21	5267728.51	-2590140.421	34613.67564	4614712.748
Cp [kJ/kmol-K]	86.71694675	215.845671	113.7470913	154.5506192	242.9797235
Mole Fraction					
METHANE	2.36E-13	5.90E-43	2.75E-13	5.60E-26	5.79E-51
ETHANE	0.008286456	1.11E-10	0.009662067	2.01E-09	4.21E-18
PROPANE	0.84205608	0.0009168	0.979999945	0.005896984	5.66E-08
n-BUTANE	0.062178808	0.21681188	0.000290526	0.5075339	0.029088215
ISOBUTANE	0.087284303	0.15288067	0.010047456	0.472466127	0.00347946
ISOPENTANE	0.000170576	0.2303886	5.10E-11	0.01220486	0.347946862
n-PENTANE	2.38E-05	0.11459938	1.88E-12	0.001897996	0.176732583
n-HEXANE	3.34E-12	0.07896595	2.97E-26	1.30E-07	0.122932638
n-HEPTANE	1.90E-17	0.07075507	1.21E-36	6.94E-11	0.110150217
n-OCTANE	1.83E-23	0.05122104	6.21E-48	2.14E-14	0.079739984
n-NONANE	2.92E-29	0.0346661	6.93E-59	7.92E-18	0.053967555
n-DECANE	3.45E-35	0.02031453	1.31E-70	3.50E-21	0.031625288
n-UNDECANE	2.62E-42	0.00988122	1.84E-82	5.73E-25	0.015382909
n-DODECANE	1.52E-47	0.00919195	1.40E-92	4.25E-28	0.014309864
n-TRIDECANE	9.54E-96	0.00482157	1.00E-100	1.38E-31	0.007506127
n-TETRADECANE	1.00E-100	0.00458526	1.00E-100	8.25E-35	0.007138243
CARBON DIOXIDE	4.82E-09	9.74E-32	5.62E-09	7.84E-19	9.86E-40
NITROGEN	2.32E-25	2.64E-65	2.70E-25	7.38E-42	3.12E-73

	Ports	ConnectedTo		
Ene In	EnergyFeed_36_reboilerQ	Energy [kJ/h]	11558805.69	=
Ene Out	EnergyFeed_0_condenserQ	Energy [kJ/h]	26542107.71	=
Signal	Variable_0_Reflux	Generic	1.321662159	=
	Variable_19_MoleFraction	Fraction	0.98	*
	Variable_0_MoleFraction	Fraction	0.98	*
	Variable_36_MoleRatio	Fraction	0.01	*
	DegSubCool_0_dsc	DT [C]	0	*

Tower Profiles

Tower	Stage	Feed/Draw	EnergyFeed	T	P	Liquid Flow	Vapour Flow
		kgmole/h	kJ/h	C	kPa	kgmole/h	kgmole/h
LQ*	1 (condens	L*=695.83	Q=-2.6542e+	4.181559845	550	919.6568374	2.57E-40
	2			5.494900906	551.3157895	915.8567111	1615.4904
	3			6.387790695	552.6315789	908.0355165	1611.6902
	4			7.472953577	553.9473684	897.5245081	1603.869
	5			8.858843366	555.2631579	885.1318928	1593.358
	6			10.51145433	556.5789474	872.0640748	1580.9654
	7			12.31139306	557.8947368	859.611423	1567.8976
	8			14.10688878	559.2105263	848.6259743	1555.4449

	9			15.76116634	560.5263158	839.3216101	1544.4595
	10			17.21877135	561.8421053	831.4424251	1535.1551
	11			18.47506607	563.1578947	824.5413949	1527.2759
F	12	F=811.35		19.5672344	564.4736842	784.1771694	1520.3749
	13			24.8135907	565.7894737	773.2413757	668.66373
	14			30.71254315	567.1052632	767.4182708	657.72794
	15			36.15462815	568.4210526	766.690901	651.90483
	16			40.42108025	569.7368421	768.0863444	651.17746
	17			43.48305431	571.0526316	769.0503374	652.57291
	18			45.69126897	572.3684211	768.4685872	653.5369
	19			47.45991634	573.6842105	765.9214911	652.95515
L*	20	L*=295.99		49.17332427	575	464.6927357	650.40805
	21			51.28255751	576.4705882	458.6223157	645.16686
	22			53.29431329	577.9411765	451.5609617	639.09644
	23			55.52852653	579.4117647	443.2658304	632.03509
	24			58.16315544	580.8823529	434.0754391	623.73996
	25			61.19804528	582.3529412	424.8439814	614.54957
	26			64.45769366	583.8235294	416.1163901	605.31811
	27			67.76530611	585.2941176	405.8629363	596.59052
	28			71.53280599	586.7647059	374.3134735	586.33706
F	29	F=504.61		81.0830856	588.2352941	775.9249946	554.7876
	30			83.04066508	589.7058824	778.0576171	451.78707
	31			85.37753273	591.1764706	780.8958633	453.9197
	32			88.0783677	592.6470588	784.7576033	456.75794
	33			91.04652443	594.1176471	789.6328091	460.61968
	34			94.15778165	595.5882353	794.5664842	465.49489
	35			97.47963064	597.0588235	795.2721142	470.42856
	36			102.1567942	598.5294118	768.0689029	471.13419
LQ*	37 (reboiler	L*=324.14	Q=1.1559e+0	116.1432183	600	324.1379207	443.93098

HeatExchangerUA: /E-1

Ports

		In0	In1	Out0	Out1
Connected To		/441.Out	/CONDENSA	/S4.In	/C5+_PROD.In
VapFrac		0	2.38E-08	0.017599125	0
T [C]		25.83729665	116.143218	36.85	63.07978087
P [kPa]		3009	600	2936	590
MoleFlow [kgmole/h]		2483.505681	324.137921	2483.505681	324.1379207
MassFlow [kg/h]		100498.0305	29026.7902	100498.0305	29026.7902
Energy [kJ/h]		-771563.5895	4614712.75	3149090.564	694058.5945

DeltaT1	DT [C]	53.06344
DeltaT0	DT [C]	-11.0127
EnergyLost	Energy [kJ/h]	-3920654
DeltaP1	DP [Kpa]	10
DeltaP0	DP [Kpa]	73
UA0_1	UA [W/K]	19571.91
DeltaT0_1_	DT [C]	-37.24248
TApproach_	DT [C]	37.24248
DeltaT0_1_	DT [C]	-79.29322

Profiles-1

Segment	UA [W/K]	LMTD [C]	T [C]	T [C]	EnergyAcum [kJ/h]	EnergyAcum [kJ/h]
	Side(0,1)	Side(0,1)	Side0	Side1	Side0	Side1
0			25.8372967	63.07978087	0	0
1	19571.913	-55.64456518	36.85	116.1432183	3920654.153	3920654.153

HeatExchangerUA: /E-2

Ports

		In0	In1	Out0	Out1
Connected To		/S9.Out	/501.Out	/PROPANE_PR	/SALE_GAS.In
VapFrac		0	1	0	1
T [C]		4.963052095	69.560607	20	67.17588377
P [kPa]		1450	3467	1440	3430
MoleFlow [kgmole/h]		695.8335236	13223.8204	695.8335236	13223.82041
MassFlow [kg/h]		30689.80768	214796.356	30689.80768	214796.3558
Energy [kJ/h]		-2523583.629	148362162	-1300737.96	147139316.7
Cp [kJ/kmol-K]		113.085055	40.9956477	121.0260974	40.91583314

DeltaT1	DT [C]	2.384723
DeltaT0	DT [C]	-15.03695
EnergyLost_	Energy [kJ/h]	-1222846
DeltaP1	DP [Kpa]	37
DeltaP0	DP [Kpa]	10
UA0_1	UA [W/K]	6104.159
DeltaT0_1_	DT [C]	-62.21283
TApproach_	DT [C]	49.56061
DeltaT0_1_	DT [C]	-49.56061

Profiles-1

Segment	UA [W/K]	LMTD [C]	T [C]	T [C]	EnergyAcum [kJ/h]	EnergyAcum [kJ/h]
	Side(0,1)	Side(0,1)	Side0	Side1	Side0	Side1
0			4.96305209	67.17588377	0	0
1	6104.1586	-55.64720292	20	69.56060704	1222845.669	1222845.669

HeatExchangerUA: /E-3

Ports

		In0	In1	Out0	Out1
Connected To		/425N.Out	/BUTANE.Ou	/S11.In	/BUTANE_PROD.In
VapFrac		1	0	1	0
T [C]		17	49.1733243	19.38439151	21
P [kPa]		2203	575	2203	565
MoleFlow [kgmole/h]		13223.82041	295.987566	13223.82041	295.9875663
MassFlow [kg/h]		214796.3558	17237.5179	214796.3558	17237.51785
Energy [kJ/h]		123355379.2	34613.6756	124581129.5	-1191136.587
Cp [kJ/kmol-K]		38.85958327	154.550619	38.89032172	140.1453129

DeltaT1	DT [C]	28.17332
DeltaT0	DT [C]	-2.384392
EnergyLost_	Energy [kJ/h]	-1225750
DeltaP1	DP [Kpa]	10
DeltaP0	DP [Kpa]	0
UA0_1	UA [W/K]	26509.15
DeltaT0_1_	DT [C]	-4
TApproach_	DT [C]	4
DeltaT0_1_	DT [C]	-29.78893

Profiles-1

Segment	UA [W/K]	LMTD [C]	T [C]	T [C]	EnergyAcum [kJ/h]	EnergyAcum [kJ/h]
	Side(0,1)	Side(0,1)	Side0	Side1	Side0	Side1
0			17	21	0	0
1	26509.149	-12.84410094	19.3843915	49.17332427	1225750.262	1225750.262

HeatExchangerUA: /E-5

Ports

		In0	In1	Out0	Out1
Connected To		/700N.Out	/401B.Out	/700B.In	/S26.In
VapFrac		1	0.85743808	1	0.841060898
T [C]		-50.0000005	-33.5	-35.3908149	-36.5
P [kPa]		5380	5460	5370	5445
MoleFlow [kgmole/h]		4411.986853	15690	4411.986853	15690
MassFlow [kg/h]		77577.1062	316711.165	77577.1062	316711.165
Energy [kJ/h]		22539679.41	74591339.9	26773047.77	70357971.57
Cp [kJ/kmol-K]		75.89511945	65.7238857	58.31329021	67.0507135

DeltaT1	DT [C]	3
DeltaT0	DT [C]	-14.60919
EnergyLost_	Energy [kJ/h]	-4233368
DeltaP1	DP [Kpa]	15
DeltaP0	DP [Kpa]	10
UA0_1	UA [W/K]	199110.9
DeltaT0_1_	DT [C]	-13.5
TApproach_	DT [C]	1.890815
DeltaT0_1_(	DT [C]	-1.890815

Profiles-1

Segment	UA [W/K]	LMTD [C]	T [C]	T [C]	EnergyAcum [kJ/h]	EnergyAcum [kJ/h]
	Side(0,1)	Side(0,1)	Side0	Side1	Side0	Side1
0			-50.0000005	-36.5	0	0
1	199110.9	-5.905933229	-35.3908149	-33.5	4233368.356	4233368.356

MultiSidedHeatExchangerOp: /E-4

Ports

		In0	In1	In2	Out0	Out1	Out2
Connected To		/703.Out	/700C.Out	/S14.Out	/S10.In	/S24.In	/S12.In
VapFrac		1	0	0.87459222	1	0	0.8294001
T [C]		-101.3703492	-74	-85.5	-78.43277995	-99.5	-89
P [kPa]		2250	5360	2370	2240	5360	2370
MoleFlow [kgmole/h]		13223.82041	4411.98685	8193.68987	13223.82041	4411.986853	8193.6899
MassFlow [kg/h]		214796.3558	77577.1062	144071.7686	214796.3558	77577.1062	144071.77
Energy [kJ/h]		58946452.48	3942427.81	32438284.24	73237007.95	-6577687.579	28667844
Cp [kJ/kmol-K]		53.31805765	140.383082	50.76883277	43.04651858	73.13461547	52.76698

CompositeAppT_AppT		DT [C]	0.547894591
DeltaP1		DP [Kpa]	0
DeltaP0		DP [Kpa]	10
DeltaP2		DP [Kpa]	0
DeltaT0_1_In		DT [C]	-1.870349175
DeltaT0_2_In		DT [C]	-12.37034918
UA1_2			
DeltaT1		DT [C]	25.5
DeltaT0		DT [C]	-22.93756922
DeltaT2		DT [C]	3.5
UA0_2			
OverallUA_Var		UA [W/K]	1751548.904
UA0_1			
DeltaT1_2_Out		DT [C]	11.5
DeltaT1_2_In		DT [C]	-10.5
DeltaT0_1_Out		DT [C]	-4.432779953

Profiles-1

Segment	UA [W/K]	LMTD [C]	T [C]	T [C]	T [C]	EnergyAcum [kJ/h]	EnergyAcun	EnergyAcum [kJ/h]
	Side(0,1)	Side(0,1)	Side0	Side1	Side2	Side0	Side1	Side2
0			-101.370349	-99.5	-89	0	0	0
1	652143.81	-5.050913379	-78.43278	-74	-85.5	14290555.47	10520115	3770440.078
							3969.5987	2922.254275

MultiSidedHeatExchangerOp: /E-12

Ports

		In0	In1	In2	In3	In4
Connected To		/FEED_311.Out	/703A.Out	/PA1.Out	/ETHANE.Out	/QC_IN.Out
VapFrac		0.999802013	1	1.67E-09	0.288664453	0
T [C]		24.8	-41.4270304	-1.484017761	-27.83030011	-12
P [kPa]		5480	2240	2386	1200	550
MoleFlow [kgmole/h]		15690	13223.8204	4244	1376.746465	16900
MassFlow [kg/h]		316711.165	214796.356	155769.7972	41752.22571	745376.178
Energy [kJ/h]		142801675.3	93192974.1	-9782659.883	-407655.5966	-93024307.37
Cp [kJ/kmol-K]		53.22753689	39.4205112	107.7261178	82.49116838	106.7856859

VapFrac		0.877912449	1	0.132720766	1	0
T [C]		-29.31906844	17	6	20	-6.927554266
P [kPa]		5470	2203	2376	1190	500
MoleFlow [kgmole/h]		15690	13223.8204	4244	1376.746465	16900
MassFlow [kg/h]		316711.165	214796.356	155769.7972	41752.22571	745376.178
Energy [kJ/h]		80384737.37	123355379	-1698055.907	14538175.68	-83800209.88
Cp [kJ/kmol-K]		64.01669831	38.8595833	106.0807526	56.88693458	108.8378406

DeltaT0_4_In	DT [C]	31.7275543
DeltaT1_2_Out	DT [C]	-39.9430126
DeltaT2_4_In	DT [C]	12.9275543
DeltaT0_2_Out	DT [C]	-27.8350507
DeltaT2_4_Out	DT [C]	10.5159822
DeltaT3_4_Out	DT [C]	-15.8303001
DeltaT0_4_Out	DT [C]	-17.3190684
DeltaT1_3_In	DT [C]	-3
DeltaT1_4_In	DT [C]	23.9275543
DeltaT2_3_Out	DT [C]	26.3462824
CompositeAppT_AppT	DT [C]	0.14175108
DeltaT1_2_In	DT [C]	11
DeltaT0_1_Out	DT [C]	12.1079619
DeltaP4	DP [Kpa]	50
DeltaT3_4_In	DT [C]	26.9275543
DeltaP1	DP [Kpa]	37

DeltaP0		DP [Kpa]	10
DeltaP3		DP [Kpa]	10
DeltaP2		DP [Kpa]	10
DeltaT0_1_In		DT [C]	7.8
DeltaT1_3_Out		DT [C]	-13.59673025
DeltaT2_3_In		DT [C]	-14
DeltaT3		DT [C]	-47.83030011
DeltaT1		DT [C]	-58.42703036
DeltaT0		DT [C]	54.11906844
DeltaT2		DT [C]	-7.484017761
DeltaT0_2_In		DT [C]	18.8
DeltaT0_3_In		DT [C]	4.8
DeltaT1_4_Out		DT [C]	-29.42703036
DeltaT0_3_Out		DT [C]	-1.488768327
OverallUA_Var		UA [W/K]	6382651.982
DeltaT4		DT [C]	-5.072445734

Profiles-1

T [C]	T [C]	T [C]	T [C]	T [C]	EnergyAcum [kJ/h]	EnergyAcum [kJ/h]	EnergyAcum	EnergyAcum [k	EnergyAc
Side0	Side1	Side2	Side3	Side4	Side0	Side1	Side2	Side3	Side4
2.48E+01	1.70E+01	6.00E+00	20	-6.927554266	0	0	0	0	0
-29.31907	-41.42703	-1.484017761	-27.8303001	-12	62416937.89	30162405.15	8084604	14945831.28	9224097
					17338.0383	8378.445875	2245.7233	4151.6198	2562.249

MultiSidedHeatExchangerOp: /E-13

Ports

		In0	In1	In2	In3	In4	In5
Connected To		/401A.Out	/PA5.Out	/PA3.Out	/420AN.Out	/403N.Out	/700B.Out
VapFrac		0.877912449	0	1.19E-14	1	0.295440897	1
T [C]		-29.31906844	-91.6673716	-73.66876557	-50.0000005	-53.9337108	-35.39081
P [kPa]		5470	2374	2378	5380	2468	5370

MassFlow [kg/h]	316711.165	97952.1495	77899.46819	144071.7686	79923.78357	77577.106
Energy [kJ/h]	80384737.37	-14229200.1	-13156081.37	41859404.62	-7383497.197	26773048
Cp [kJ/kmol-K]	64.01669831	72.3540333	75.00547474	75.89511945	75.35649443	58.31329

	Out0	Out1	Out2	Out3	Out4	Out5
Connected To	/401B.In	/PA6.In	/PA4.In	/420BN.In	/403.In	/700C.In
VapFrac	0.857438077	0.17973416	0.284687873	1	0.382997423	0
T [C]	-33.5	-89	-55.26	-30	-40	-74
P [kPa]	5460	2374	2378	5370	2468	5360
MoleFlow [kgmole/h]	15690	4600	2975	8193.68987	2493.75452	4411.9869
MassFlow [kg/h]	316711.165	97952.1495	77899.46819	144071.7686	79923.78357	77577.106
Energy [kJ/h]	74591339.92	-9037434.88	-4205332.027	52228348.63	-3270938.401	3942427.8
Cp [kJ/kmol-K]	65.72388567	67.9237121	71.16257572	55.35654874	75.46989056	140.38308

Ports	ConnectedTo			
CompositeAppT_AppT	DT [C]	0.68093156	=	
DeltaT1_2_Out	DT [C]	-17.998606	=	
DeltaT2_4_Out	DT [C]	-19.7350548	=	
DeltaT1_5_In	DT [C]	-15	=	
DeltaT1_4_In	DT [C]	-49	=	
DeltaT0_4_Out	DT [C]	20.4337108	=	
DeltaT0_5_Out	DT [C]	1.8908149	=	
DeltaT4_5_Out	DT [C]	-18.5428959	=	
DeltaT2_3_Out	DT [C]	-23.6687651	=	
DeltaT0_1_Out	DT [C]	58.1673716	=	
DeltaT2_5_Out	DT [C]	-38.2779507	=	
DeltaT0	DT [C]	4.18093156	=	
DeltaT2_3_In	DT [C]	-25.26	=	
DeltaT3_5_Out	DT [C]	-14.6091856	=	
DeltaT1_5_Out	DT [C]	-56.2765567	=	
DeltaT3_5_In	DT [C]	44	=	

DeltaT0_4_In		DT [C]	10.68093156
DeltaT0_2_In		DT [C]	25.94093156
DeltaP4		DP [Kpa]	0
DeltaT3_4_Out		DT [C]	3.933710306
DeltaT1_3_In		DT [C]	-59
DeltaP1		DP [Kpa]	0
DeltaT1_2_In		DT [C]	-33.74
DeltaP5		DP [Kpa]	10
DeltaT0_2_Out		DT [C]	40.16876557
DeltaT3_4_In		DT [C]	10
DeltaP0		DP [Kpa]	10
DeltaP3		DP [Kpa]	10
DeltaP2		DP [Kpa]	0
DeltaT0_1_In		DT [C]	59.68093156
DeltaT1		DT [C]	-2.667371571
DeltaT2_4_In		DT [C]	-15.26
DeltaT0_3_In		DT [C]	0.680931562

DeltaT1_4_Out	DT [C]	-37.7336608	=
DeltaT0_3_Out	DT [C]	16.5000005	=
OverallUA_Var	UA [W/K]	1747344.08	=
DeltaT4	DT [C]	-13.9337108	=
DeltaT4_5_In	DT [C]	34	=

DeltaT2		DT [C]	-18.40876557
DeltaT5		DT [C]	38.6091851
DeltaT2_5_In		DT [C]	18.74
DeltaT0_5_In		DT [C]	44.68093156
DeltaT3		DT [C]	-20.0000005

Profiles-1

Segment	T [C]	T [C]	T [C]	T [C]	T [C]	T [C]
	Side0	Side1	Side2	Side3	Side4	Side5
0	-29.31906844	-89	-55.26	-30	-40	-74
1	-33.5	-91.6673716	-73.66876557	-50.0000005	-53.9337108	-35.39081

EnergyAcum [kJ/	EnergyAcum	EnergyAcum [k	EnergyAcum [kJ/	EnergyAcum [kJ/h]	EnergyAcum [kJ/h]
Side0	Side1	Side2	Side3	Side4	Side5
0	0	0	0	0	0
5793397.453	5191765.27	8950749.341	10368944	4112558.796	22830620
1609.27707	1442.15702	2486.319261	2880.262223	1142.377443	6341.8389

MultiSidedHeatExchangerOp: /E-14

Ports		In0	In1	In2	In3	Out0	Out1	Out2	Out3
Connected To		/S10.Out	/413A_N.Out	/410N.Out	/421A_N.Out	/703A.In	/413N.In	/411A.In	/S13.In
VapFrac		1	0.33019428	1	0.974240648	1	0.2654213	0.955247205	0.943505
T [C]		-78.43277995	-70.7838712	-36.50000017	-68.01130038	-41.42703036	-75	-50	-76
P [kPa]		2240	2575	5445	2375	2240	2565	5380	2370
MoleFlow [kgmole/h]		13223.82041	590.568757	13196.24548	8193.68987	13223.82041	590.56876	13196.24548	8193.69
MassFlow [kg/h]		214796.3558	15138.5066	236787.3814	144071.7686	214796.3558	15138.507	236787.3814	144071.8
Energy [kJ/h]		73237007.95	-1050338.67	77741467.93	45442101.68	93192974.08	-1429839	63348746.06	40258358
Cp [kJ/kmol-K]		43.04651858	67.0632683	62.85380448	45.38432782	39.4205112	68.083496	76.34004558	47.25453

DeltaT1_2_Out	DT [C]	-34.283871	
DeltaT0_2_In	DT [C]	-28.43278	
DeltaT1_3_In	DT [C]	1	
DeltaT2_3_Out	DT [C]	31.5113002	
DeltaT1_2_In	DT [C]	-25	
DeltaT0_1_Out	DT [C]	29.3568408	

DeltaP2		DP [Kpa]	65
CompositeAppT_AppT		DT [C]	0.105164991
DeltaT1_3_Out		DT [C]	-2.772570786
DeltaT2_3_In		DT [C]	26
DeltaT3		DT [C]	7.988699618
DeltaT1		DT [C]	4.216128832

DeltaT0		DT [C]	-37.0057496
DeltaT0_2_Out		DT [C]	-4.92703019
DeltaP1		DP [Kpa]	10
DeltaP0		DP [Kpa]	0
DeltaP3		DP [Kpa]	5

DeltaT0_1_In		DT [C]	-3.432779953
DeltaT0_3_In		DT [C]	-2.432779953
DeltaT2		DT [C]	13.49999983
DeltaT0_3_Out		DT [C]	26.58427002
OverallUA_Var		UA [W/K]	2780008.805

Segment	T [C]	T [C]	T [C]	T [C]	EnergyAcum [kJ/h]	EnergyAcum [kJ/h]	EnergyAcum [kJ/h]	EnergyAcum [kJ/h]
	Side0	Side1	Side2	Side3	Side0	Side1	Side2	Side3
0	-78.43278	-7.50E+01	-50	-76	0	0	0	0
1	-41.42703	-70.78387117	-36.5000002	-68.01130038	19955966.13	379500.5904	14392722	5183743.68
					5543.323926	105.4168307	3997.9783	1439.9288

Splitter: /SP-1

Ports		In	Out0	Out1
Connected To		/440.Out	/441.In	/441T.In
VapFrac		0	0	0
T [C]		25.83729665	25.8372967	25.83729665
P [kPa]		3009	3009	3009
MoleFlow [kgmole/h]		2658.505681	2483.50568	175
MassFlow [kg/h]		107579.615	100498.031	7081.584499
Energy [kJ/h]		-825931.7471	-771563.59	-54368.15755
Cp [kJ/kmol-K]		124.2754239	124.275424	124.2754239

Ports	ConnectedTo		
FlowFraction1		Fraction	0.065826
FlowFraction0		Fraction	0.934174

Splitter: /SP-2

Ports

	In	Out0	Out1
Connected	/420N.Out	/420AN.In	/700N.In
VapFrac	1	1	1
T [C]	-50.0000005	-50.0000005	-50.0000005
P [kPa]	5380	5380	5380
MoleFlow [k	12605.67672	8193.68987	4411.986853
MassFlow [	221648.8748	144071.769	77577.1062
Energy [kJ/	64399084.04	41859404.6	22539679.41
Cp [kJ/kmo	75.89511945	75.8951194	75.89511945

Ports	ConnectedTo		
FlowFraction1		Fraction	0.35
FlowFraction0		Fraction	0.65

ReboiledAbsorber: /DEMETHANSIER

Ports		Feed_0_overhea	Feed_15_fee	Feed_19_feed	Feed_4_feed	Feed_6_feed	Feed_9_fee	Feed_9_feed_1
Connected To		/704.Out	/FROM_A451	/PA2.Out	/413N.Out	/PA6.Out	/403.Out	/PA4.Out
VapFrac		0.00E+00	1	0.13270579	0.265421323	0.179731289	0.3829974	0.28468413
T [C]		-94.29957548	41.91	6	-75	-89	-40	-55.26
P [kPa]		2368	2425	2376	2565	2374	2468	2378

MoleFlow [kgmole/h]	4646.36172	192.3	4244	590.5687569	4600	2493.7545	2975
MassFlow [kg/h]	92876.95146	5663.9741	155769.9941	15138.50657	97952.2219	79923.784	77899.59753
Energy [kJ/h]	-12043957.51	2166152.14	-1698686.094	-1429839.26	-9037543.575	-3270938	-4205454.529
Cp [kJ/kmol-K]	73.66982801	61.7685733	106.0815197	68.0834959	67.92377844	75.469891	71.16270731
Mole Fraction							
METHANE	7.74E-01	2.62E-01	0.055186686	0.597181084	0.717605399	0.4692741	0.440567823
ETHANE	0.175273545	0.56112344	0.599146021	0.200385717	0.206768363	0.1897869	0.423275325
PROPANE	0.037860193	0.0950199	0.188956975	0.133033375	0.054314909	0.1663502	0.096372052
n-BUTANE	0.00228096	0.01452299	0.037896621	0.023932062	0.005244601	0.0509939	0.008803574
ISOBUTANE	0.003074546	0.00894599	0.034588934	0.024180522	0.006063206	0.0436686	0.010249918
ISOPENTANE	4.21E-04	4.23E-03	0.025351593	0.010419336	0.001717156	0.0393749	0.002820112
n-PENTANE	1.65E-04	1.83E-03	0.012322089	0.004548692	0.000731149	0.019441	0.001196136
n-HEXANE	8.54E-06	1.24E-03	0.005908687	0.000776503	0.000105474	0.0099299	0.000169541
n-HEPTANE	7.75E-07	6.74E-04	0.003390409	0.000181005	2.34E-05	0.0057944	3.74E-05
n-OCTANE	1.94E-08	2.91E-04	0.000551523	1.13E-05	1.44E-06	0.0009347	2.29E-06
n-NONANE	5.52E-10	8.80E-05	0.000108647	8.29E-07	1.04E-07	0.0001823	1.65E-07
n-DECANE	1.91E-11	2.30E-05	2.99E-05	8.10E-08	1.01E-08	5.03E-05	1.60E-08
n-UNDECANE	1.00E-100	5.00E-06	2.21E-07	0	1.72E-76	0	3.04E-53
n-DODECANE	1.00E-100	2.00E-06	8.85E-08	0	8.74E-83	0	1.62E-57
n-TRIDECANE	1.00E-100	0.00E+00	1.00E-100	0	1.00E-100	0	1.00E-100
n-TETRADECANE	1.00E-100	0.00E+00	1.00E-100	0	1.00E-100	0	1.00E-100
CARBON DIOXIDE	5.45E-03	4.77E-02	0.036561207	0.003080081	0.007248451	0.0026218	0.016357854
NITROGEN	1.23E-03	2.29E-03	3.92E-07	0.00226937	0.000176351	0.0015969	0.00014784

Ports	LiquidDraw_18_L	LiquidDraw_2	LiquidDraw_5_c	LiquidDraw_8_dra	VapourDraw_0_overheadV
Connected To	/PA1.In	/C2+.In	/PA5.In	/PA3.In	/METHANE.In
VapFrac	1.67E-09	3.86E-10	0	1.19E-14	1
T [C]	-1.484017761	25	-91.66737157	-73.66876557	-94.13081382
P [kPa]	2386	2396	2374	2378	2368
MoleFlow [kgmole/h]	4244	2658.50568	4600	2975	5264.479315
MassFlow [kg/h]	155769.7972	107579.615	97952.14952	77899.46819	86023.99926
Energy [kJ/h]	-9782659.883	-1010734.66	-14229200.15	-13156081.37	24812203.85
Cp [kJ/kmol-K]	107.7261178	125.775039	72.35403328	75.00547474	51.47738866
Mole Fraction					
METHANE	0.055189727	0.00351487	0.71760628	0.440569841	0.980402902
ETHANE	0.599142557	0.53886008	0.206767686	0.423274212	0.015456467
PROPANE	0.18895675	0.25762486	0.05431477	0.096371406	0.000511487
n-BUTANE	0.037896484	0.05817845	0.005244589	0.008803518	4.54E-06
ISOBUTANE	0.034588824	0.05233322	0.006063192	0.010249852	1.09E-05
ISOPENTANE	0.025351493	0.04029097	0.001717152	0.002820095	1.58E-07
n-PENTANE	0.012322041	0.01966714	0.000731147	0.001196128	5.62E-08
n-HEXANE	0.005908664	0.00952436	0.000105474	0.00016954	1.65E-10

n-HEPTANE	0.003390396	0.00552566	2.34E-05	3.74E-05	7.13E-12
n-OCTANE	0.000551521	0.00090042	1.44E-06	2.29E-06	3.48E-14
n-NONANE	0.000108647	0.00017752	1.04E-07	1.65E-07	2.03E-16
n-DECANE	2.99E-05	4.89E-05	1.01E-08	1.60E-08	1.07E-18
n-UNDECANE	2.21E-07	3.62E-07	1.72E-76	3.04E-53	1.00E-100
n-DODECANE	8.85E-08	1.45E-07	8.75E-83	1.62E-57	1.00E-100
n-TRIDECANE	1.00E-100	1.00E-100	1.00E-100	1.00E-100	1.00E-100
n-TETRADECANE	1.00E-100	1.00E-100	1.00E-100	1.00E-100	1.00E-100
CARBON DIOXIDE	0.03656229	0.01328607	0.007248419	0.016357741	0.001433328
NITROGEN	3.92E-07	9.97E-10	0.000176351	0.000147841	0.002180142

Tower	Stage	Feed/Draw	EnergyFeed	T	P	Liquid Flow	Vapour Flow	L_T
		kgmole/h	kJ/h	C	kPa	kgmole/h	kgmole/h	C
VF	1	V=5264.48 F=4646.36		-94.13081382	2368	4652.45611	5264.4793	-94.13081382
	2			-94.07981026	2369.2	4652.171514	5270.5737	-94.07981026
	3			-94.01774783	2370.4	4642.422356	5270.2891	-94.01774783
	4			-93.79905676	2371.6	4577.746715	5260.54	-93.79905676
F	5	F=590.57		-92.47231945	2372.8	4887.159496	5195.8643	-92.47231945
L*	6	L*=4600.00		-91.66737157	2374	124.2644451	4914.7083	-91.66737157
F	7	F=4600.00		-88.01818247	2375.333333	3779.219733	4751.8133	-88.01818247
	8			-84.22228953	2376.666667	3478.956826	3806.7686	-84.22228953
L*	9	L*=2975.00		-73.66876557	2378	25.32977147	3506.5057	-73.66876557
FF	10	F=2493.75 F=2975.00		-49.44813592	2378.8	3597.727821	3027.8786	-49.44813592
	11			-49.31056424	2379.6	3599.484111	1131.5221	-49.31056424
	12			-49.00178619	2380.4	3601.495563	1133.2784	-49.00178619
	13			-48.20807708	2381.2	3606.136598	1135.2899	-48.20807708
	1.40E+01			-46.19789168	2382	3619.651271	1139.9309	-46.19789168
	15			-41.48691586	2382.8	3660.44406	1153.4456	-41.48691586
F	16	F=192.30		-32.02383151	2383.6	3735.137118	1194.2384	-32.02383151
	17			-23.66635303	2384.4	3873.241386	1076.6314	-23.66635303
	18			-12.468138	2385.2	4099.829797	1214.7357	-12.468138
L*	19	L*=4244.00		-1.484017761	2386	101.8582803	1441.3241	-1.484017761
F	20	F=4244.00		8.5930146	2388.5	3860.842825	1687.3526	8.5930146
	21			12.13423007	2391	3950.344721	1202.3371	12.13423007
	22			15.91724264	2393.5	3977.810246	1291.839	15.91724264
LQ*	23 (reboiler	L=2658.51	Q=1.6154e+0	25	2396	2658.505681	1319.3046	25

SpecialProps:
/Prop-1

	In	Out				
Connected To	/SALE_GAS.(					
VapFrac	1	1		Ports	ConnectedTo	
T [C]	67.1758838	67.17588377	Signal	GHVStdGasVol		StdGasVolH [k 37385.24
P [kPa]	3.43E+03	3.43E+03		WaterContent		StdGasVolMas 0
MoleFlow [kgmol	13223.82041	13223.82041		CO2_Conte		Fraction 0.000506

MassFlow [kg/h]	214796.356	214796.3558		DewPoint		T [C]	-90.63837
Energy [kJ/h]	147139317	147139316.7		H2S_Content		StdGasVolTrac	0
Cp [kJ/kmol-K]	40.9158331	40.91583314		NHVStdGasVol		StdGasVolH [kJ	33697.36
Mole Fraction							
METHANE	0.9850471	0.985047103					
ETHANE	0.00512503	0.005125033					
PROPANE	0.00017914	0.000179141					
n-BUTANE	1.57E-06	1.57E-06					
ISOBUTANE	4.01E-06	4.01E-06					
ISOPENTANE	5.05E-08	5.05E-08					
n-PENTANE	1.96E-08	1.96E-08					
n-HEXANE	1.93E-10	1.93E-10					
n-HEPTANE	2.68E-12	2.68E-12					
n-OCTANE	1.24E-14	1.24E-14					
n-NONANE	7.76E-17	7.76E-17					
n-DECANE	3.93E-19	3.93E-19					
n-UNDECANE	1.00E-100	1.00E-100					
n-DODECANE	1.00E-100	1.00E-100					
n-TRIDECANE	1.00E-100	1.00E-100					
n-TETRADECAN	1.00E-100	1.00E-100					
CARBON DIOXID	5.06E-04	5.06E-04					
NITROGEN	9.14E-03	9.14E-03					

SpecialProps: /Prop-2

Ports	In	Out	
Connected To	/C5+_PROD.Out		
VapFrac	0	0	=
T [C]	63.07978087	63.0797809	=
P [kPa]	590	590	=
MoleFlow [kgmole/h]	324.1379207	324.137921	=
MassFlow [kg/h]	29026.7902	29026.7902	=
Energy [kJ/h]	694058.5945	694058.595	=
Cp [kJ/kmol-K]	214.1112706	214.111271	=
Mole Fraction			
METHANE	5.79E-51	5.79E-51	=
ETHANE	4.21E-18	4.21E-18	=
PROPANE	5.66E-08	5.66E-08	=
n-BUTANE	0.029088215	0.02908822	=
ISOBUTANE	0.00347946	0.00347946	=
ISOPENTANE	0.347946862	0.34794686	=
n-PENTANE	0.176732583	0.17673258	=
n-HEXANE	0.122932638	0.12293264	=
n-HEPTANE	0.110150217	0.11015022	=

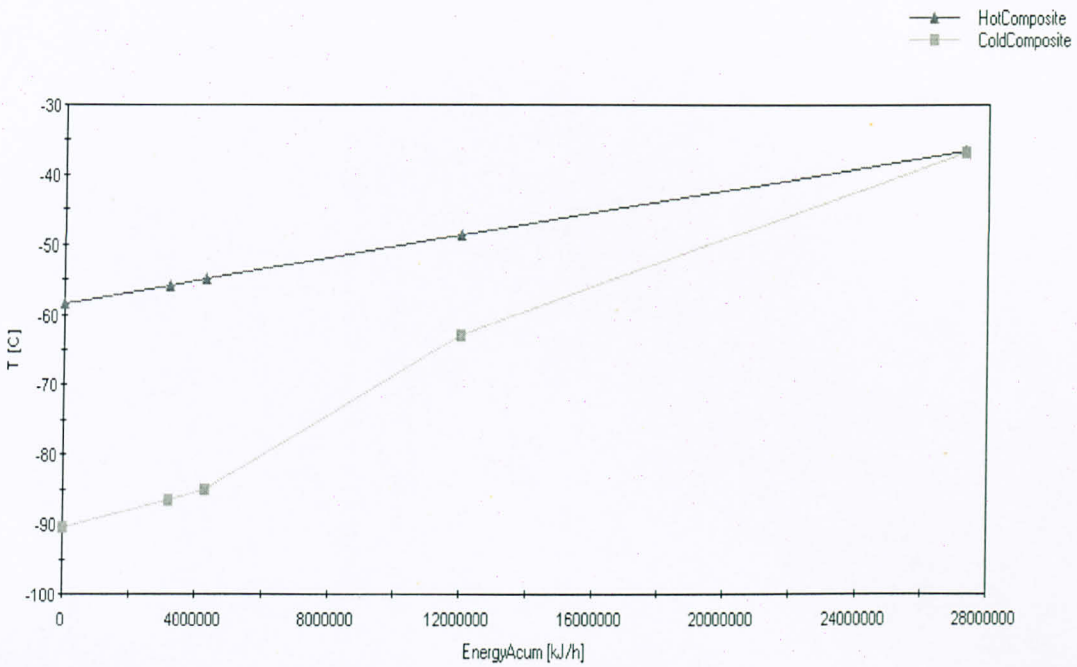
Ports	ConnectedTo		
ReidVaporPressure_D1267		P [kPa]	84.27289964
ReidVaporPressure_D323		P [kPa]	83.64916195

n-OCTANE	0.079739984	0.07973998	=
n-NONANE	0.053967555	0.05396755	=
n-DECANE	0.031625288	0.03162529	=
n-UNDECANE	0.015382909	0.01538291	=
n-DODECANE	0.014309864	0.01430986	=
n-TRIDECANE	0.007506127	0.00750613	=
n-TETRADECANE	0.007138243	0.00713824	=
CARBON DIOXIDE	9.86E-40	9.86E-40	=
NITROGEN	3.12E-73	3.12E-73	=

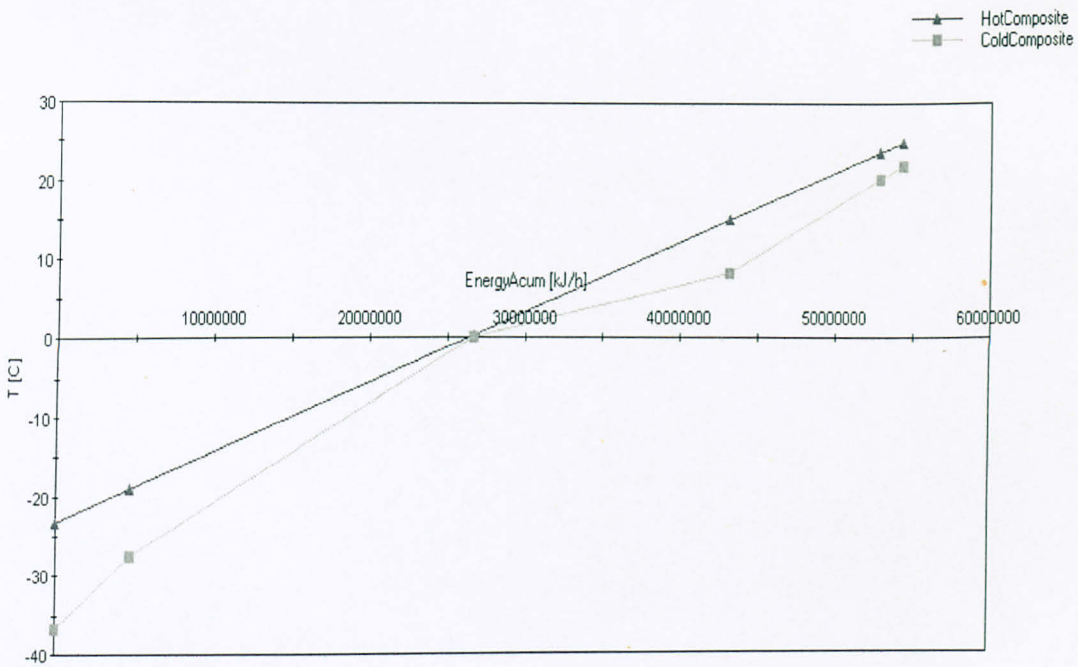
APPENDIX 6

COMPOSITE CURVES OF COLD BOXES FOR ORIGINAL GAS PLANT

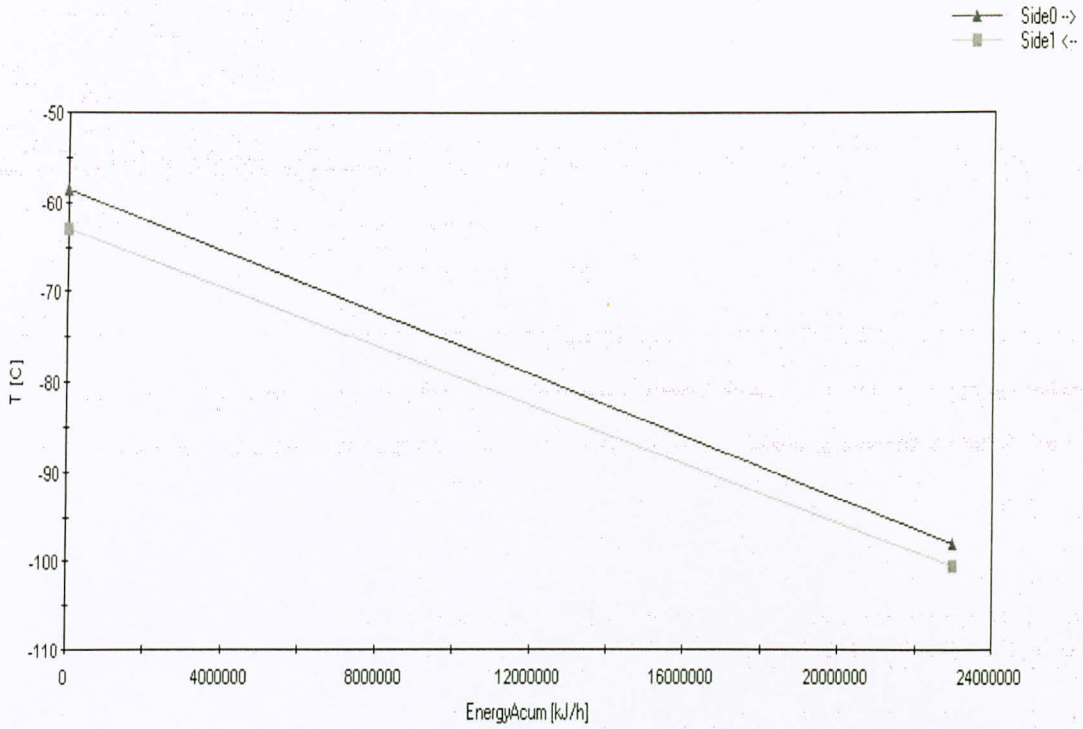
Cold box E-3



Cold box E-12



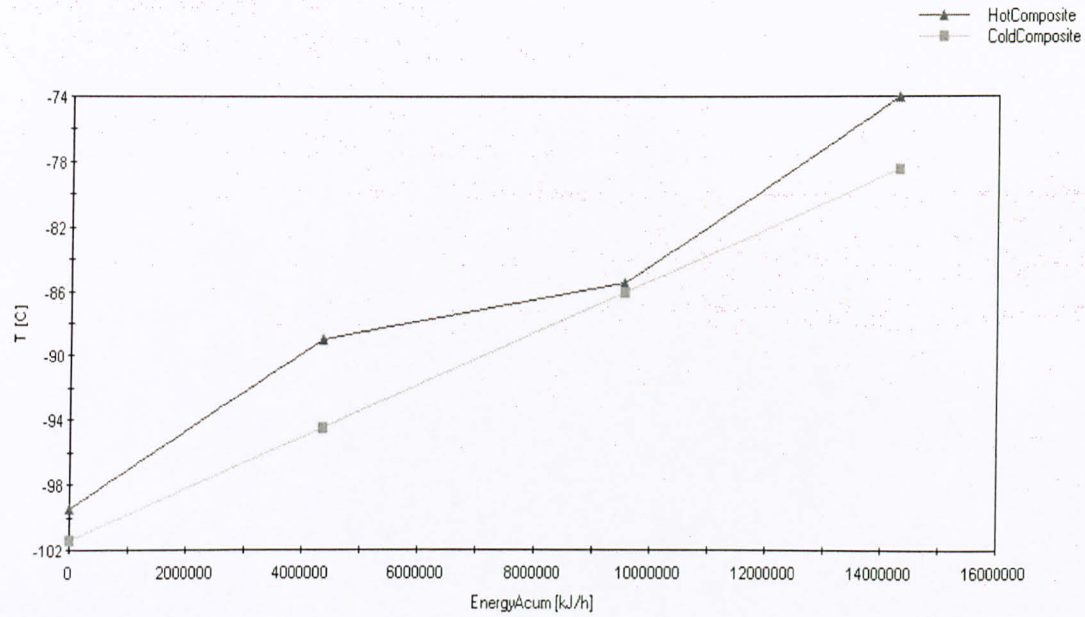
Cold box E-13


Side0 -->
Side1 <--
-50
-60
-70
-80
-90
-100
-110
T [C]
0
4000000
8000000
12000000
16000000
20000000
24000000
EnergyAcum [kJ/h]

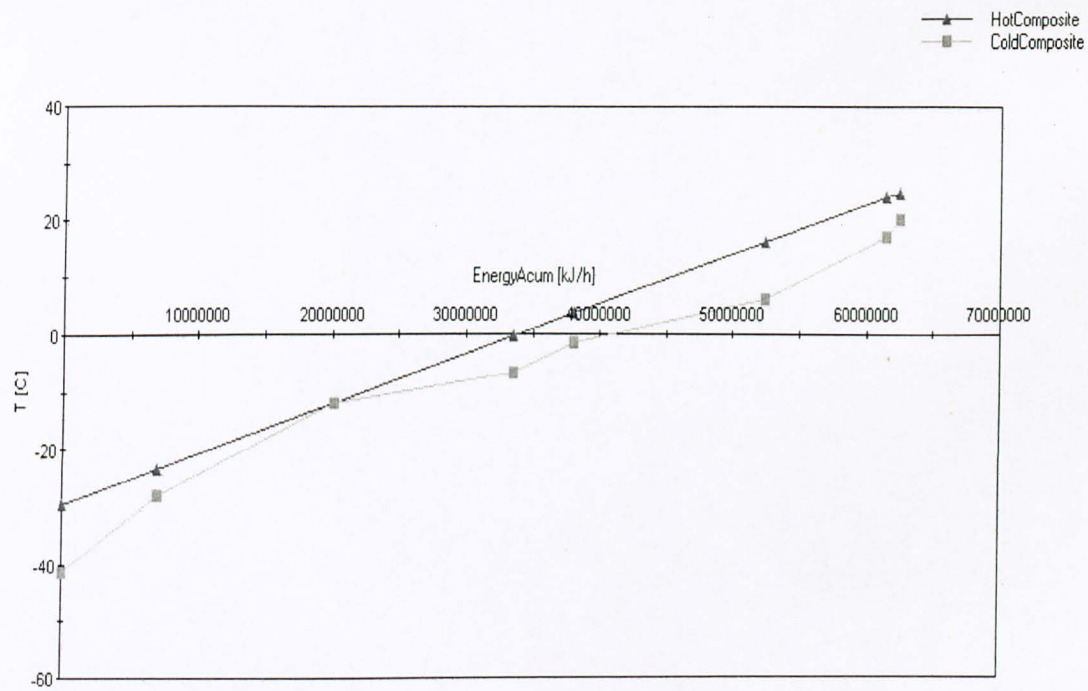
APPENDIX 7

COMPOSITE CURVES OF COLD BOXES FOR NEW GAS PLANT

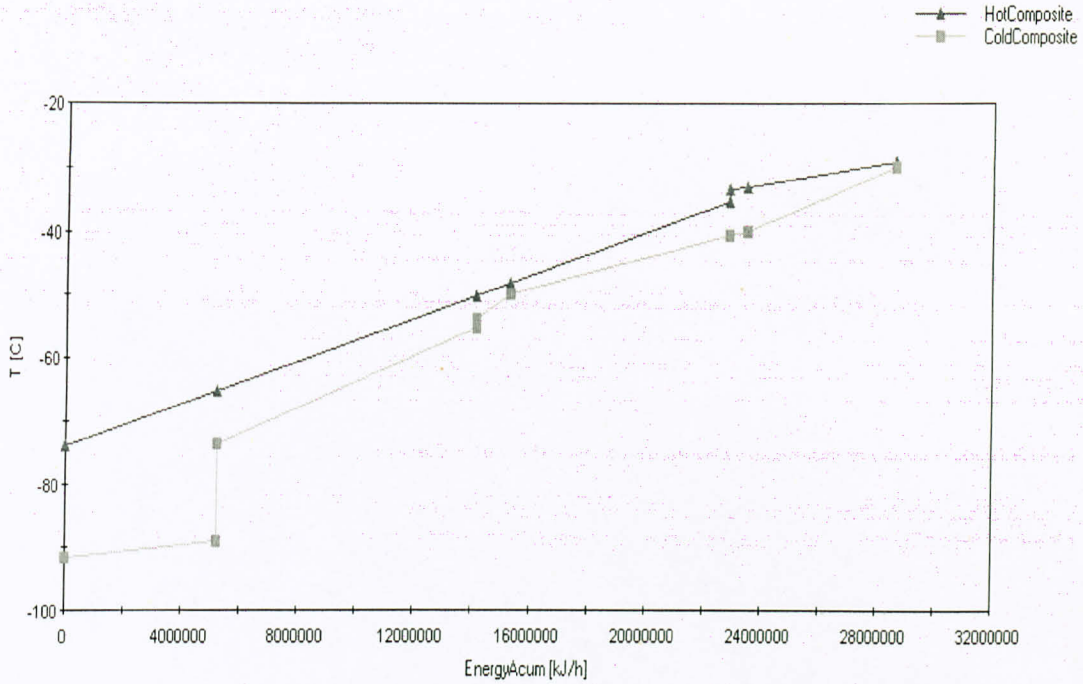
Cold box E-4



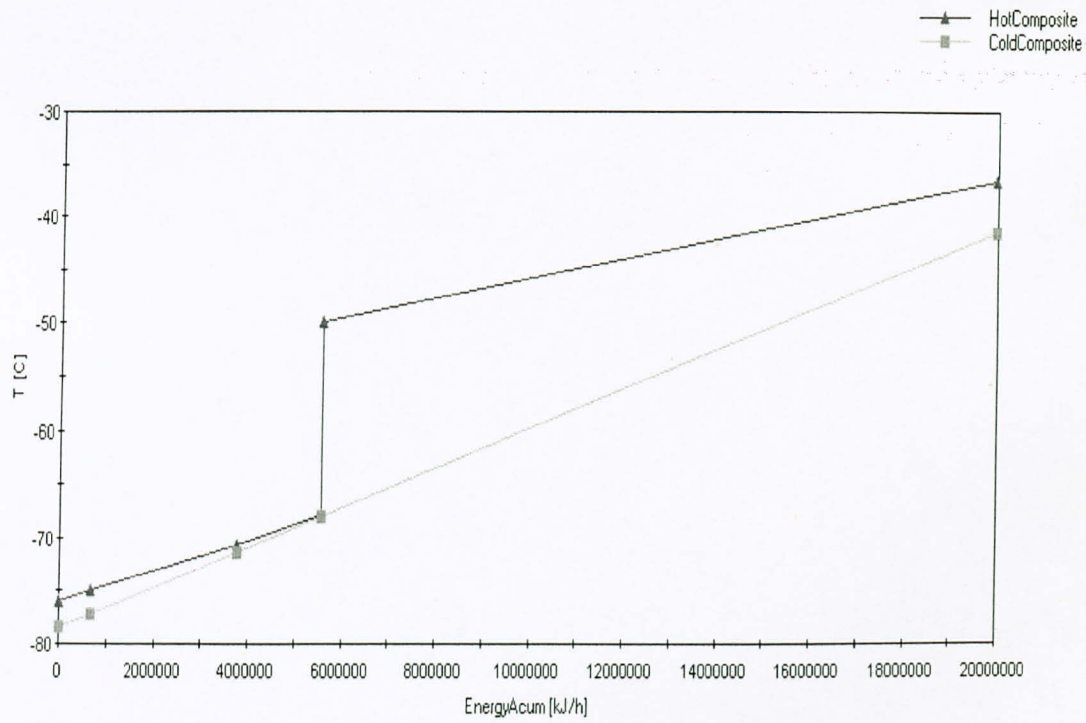
Cold box E-12



Cold box E-13



Cold box E-14



APPENDIX 8

REFRIGERATION COST

Discharge

$$T_r = \frac{T}{T_c} = \frac{330.38K}{370} = 0.893$$

$$P_r = \frac{P}{P_c} = \frac{1217K}{4260} = 0.285$$

$$B^0 = 0.083 - \frac{0.422}{T_r^{1.6}} = -0.4228$$

$$B^1 = 0.139 - \frac{0.172}{T_r^{4.2}} = -0.1377$$

$$B_r = B^0 + \omega B^1 = -0.4228 + (0.153)(-0.1377) = -0.44386$$

$$z_2 = 1 + \frac{B_r P_r}{T_r} = 0.858$$

Suction

$$T_r = \frac{T}{T_c} = \frac{310K}{370K} = 0.84$$

$$P_r = \frac{P}{P_c} = \frac{19.15K}{4260} = 0.0045$$

$Z_1 = 1$

$Z_{avg} = z_1 + z_2 = 0.929$

Gas power (actual compression power excluding mechanical losses)

Head horsepower based on polytropic compression

$$H_p = \frac{Z_{avg} R T_1}{M(n-1)/n}\left[\left(\frac{P_2}{P_1}\right)^{(n-1)/n} - 1\right]$$

which can be written in the form

$$H_p = \frac{8.314}{M}\frac{Z_{avg} T_1}{(n-1)/n}\left[\left(\frac{P_2}{P_1}\right)^{(n-1)/n} - 1\right]$$

$$H_p = \frac{8.314 \times 0.929 \times 310}{44 \times 0.3197}\left[\left(\frac{1217}{19.15}\right)^{0.3197} - 1\right]$$

$$H_p = 371.24 kJ/kg$$

Gas horsepower

$$G_p = \frac{w \times H_p}{3600 \times \eta_p} = \frac{210.14 \times 10^3 kg/h \times 371.66 kJ/kg}{0.35 \times 3600}$$

$$G_p = 61915 kW$$

$$\text{Approximate COP} = \frac{0.6 T_{EVAP}}{T_{COND} - T_{EVAP}} = \frac{0.6 \times 244}{310 - 244} = 2.22$$

$$\text{COP} = \frac{Q_c}{W} = 2.22$$

$Q_C = 2.22\ W = 2.22 G_P = 137.339 \times 10^3$ kW

Q_C = 137.339 MW

Heating value of fuel = 52000 kJ/kg
Price of fuel = RM 665/MT

Annual operating cost of refrigeration system = Fuel cost + Cooling water cost

$$= \frac{RM\,665}{MT} \times \frac{1MT}{1000kg} \times \frac{1kg}{52000kJ} \times \frac{3600s}{1hr} \times \frac{24hrs}{day} \times \frac{330days}{yr} \times 19{,}628kW$$
$$+ 2{,}001{,}435 ton/yr \times RM\,0.13/ton$$

= RM 7,417,075/yr

Total cooling duty of the refrigeration system = 19,117.4 kW

$$\text{Operating cost of refrigeration (RM/kWh)} = \frac{RM\,7{,}417{,}075/yr}{19{,}117.4kW \times 24hrs/day \times 330days/yr}$$

= RM 0.049/kWh

The operating cost of heat exchangers using refrigerant as cooling medium is calculated based on the actual load of the heat exchanger and the operating cost of refrigeration system in terms of RM/kWh

Calculation for operating cost of HE using refrigeration as cooling medium

Condenser of Prefractionator column

Q_C = 2288 kW
T_{out} = 30°C
$T_{refrigerant\ used}$ = 10°C

Shaft work for compressor $W = \frac{Q_C \times (T_H - T_C)}{0.6 \times T_C} = \frac{2288 \times (310 - 283)}{0.6 \times 283} = 364 kW$

Annual operating cost = 364kW x RM0.049/kWh x 8640 hrs/yr
= RM 141,330/yr

Condenser of Depropansier column

Q_C = 7373 kW
T_{out} = 3°C
$T_{refrigerant\ used}$ = -10°C

Shaft work for compressor $W = \frac{Q_C \times (T_H - T_C)}{0.6 \times T_C} = \frac{7373 \times (310 - 263)}{0.6 \times 263} = 2196\ kW$

Annual operating cost = 2196kW x RM0.049/kWh x 8640 hrs/yr
= RM 853,078/yr

Coolers in original designed LTSU

Q_C = 5040 kW
T_{out} = -36.5°C
$T_{refrigerant\ used}$ = -46°C

Shaft work for compressor $W = \frac{Q_C \times (T_H - T_C)}{0.6 \times T_C} = \frac{5040 \times (310 - 226)}{0.6 \times 226} = 3122\ kW$

Annual operating cost = 3122kW x RM0.049/kWh x 8640 hrs/yr
= RM 1,212,840/yr

Coolers in new designed LTSU

Q_C = 2172 kW
T_{out} = -85.5°C
$T_{refrigerant\ used}$ = -95°C

Shaft work for compressor $W = \frac{Q_C \times (T_H - T_C)}{0.6 \times T_C} = \frac{2172 \times (310 - 178)}{0.6 \times 178} = 2684\ kW$

Annual operating cost = 2684kW x RM0.049/kWh x 8640 hrs/yr
= RM 1,042,836/yr

Q_C = 2562 kW
T_{out} = -5°C
$T_{refrigerant\ used}$ = -10°C

Shaft work for compressor $W = \frac{Q_C \times (T_H - T_C)}{0.6 \times T_C} = \frac{2562 \times (310 - 263)}{0.6 \times 263} = 763\ kW$

Annual operating cost = 763kW x RM0.049/kWh x 8640 hrs/yr
= RM 296,431/yr

APPENDIX 9

REPORT OF HEAT EXCHANGER NETWORK IN LTSU AREA

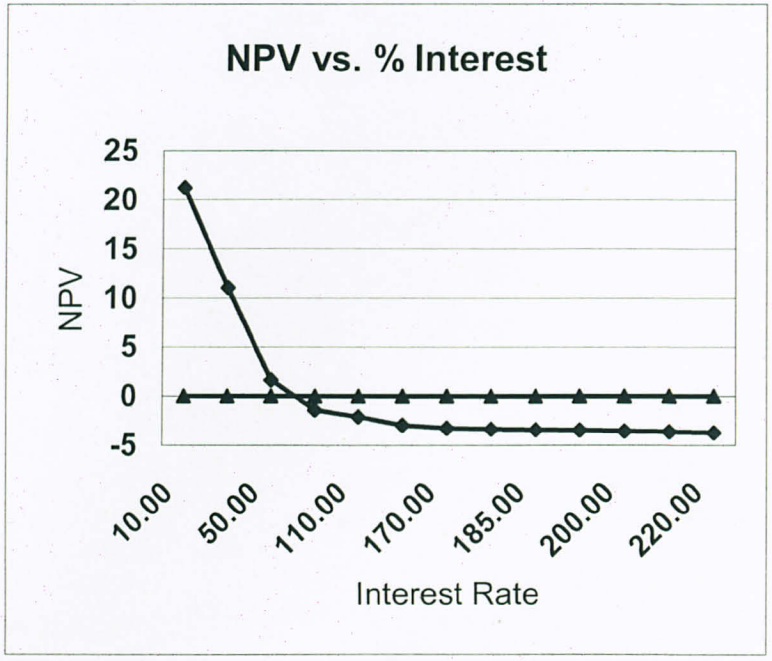
Stream	Type	TS [C]	TT [C]	DH [kW]	TTcalc [C]	dTT [C]	dDH [kW]
1 [cold utility]	Cold	-12.00	-6.93	2562.3	-6.93	0.00	0.00
2 [feed 311]	Hot	24.80	-36.50	-20123.	-36.50	0.00	0.00
3 [410 to 411]	Hot	-36.50	-50.00	-3998.0	-50.00	0.00	0.00
4 [700 to 701]	Hot	-35.39	-99.50	-9264.1	-99.50	0.00	0.00
5 [421]	Hot	-68.01	-89.00	-4659.5	-89.00	0.00	0.00
6 [413A]	Hot	-70.78	-75.00	-105.42	-75.00	0.00	0.00
7 [403A]	Cold	-53.93	-40.00	1142.4	-40.00	0.00	0.00
8 [sale gas]	Cold	-101.37	19.38	18232.	19.38	0.00	0.00
9 [ethane prod]	Cold	-27.58	20.00	4151.6	20.00	0.00	0.00
10 [PA1]	Cold	-1.48	6.00	2250.2	6.00	0.00	0.00
11 [PA3]	Cold	-73.67	-55.26	2486.3	-55.26	0.00	0.00
12 [PA5]	Cold	-91.64	-89.00	1442.2	-89.00	0.00	0.00
13 [420A]	Cold	-50.00	-30.00	2880.3	-30.00	0.00	0.00
14 [addition]	Cold	-50.00	-35.39	1175.9	-35.39	0.00	0.00
15u [refrigeratio]	Cold	-100.00	-99.00	2172.3	-99.00	0.00	0.00
16u [heating]	Hot	42.00	32.00	-345.00	32.00	0.00	0.00

NET PRESENT VALUE (NPV) AND INTERNAL RATE OF RETURN (IRR)

Description of the Project :PREFRACTIONATOR FOR PRU AREA

Year	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Savings		3.75	3.75	3.75	3.75	3.75	3.75	3.75	3.75	3.75	3.75	3.75	3.75	3.75	3.75	3.75
Maintenance		-0.27	-0.27	-0.27	-0.27	-0.27	-0.27	-0.27	-0.27	-0.27	-0.27	-0.27	-0.27	-0.27	-0.27	-0.27
Others																
Investment Cost	5.30															
Net Cash Flow		3.49	3.49	3.49	3.49	3.49	3.49	3.49	3.49	3.49	3.49	3.49	3.49	3.49	3.49	3.49

					NPV for 15 Years	
PV @ interest %	10.00	26.51	Million RM		NPV	21.21
PV @ interest %	20.00	16.29	Million RM		NPV	10.99
PV @ interest %	50.00	6.95	Million RM		NPV	1.65
PV @ interest %	90.00	3.87	Million RM		NPV	-1.43
PV @ interest %	110.00	3.17	Million RM		NPV	-2.13
PV @ interest %	150.00	2.32	Million RM		NPV	-2.98
PV @ interest %	170.00	2.05	Million RM		NPV	-3.25
PV @ interest %	180.00	1.94	Million RM		NPV	-3.36
PV @ interest %	185.00	1.88	Million RM		NPV	-3.42
PV @ interest %	190.00	1.83	Million RM		NPV	-3.47
PV @ interest %	200.00	1.74	Million RM		NPV	-3.56
PV @ interest %	210.00	1.66	Million RM		NPV	-3.64
PV @ interest %	220.00	1.58	Million RM		NPV	-3.72



Notes:

1. Figures are in Million RM
2. Assumed planned life of invesment is 15 years
3. Investment Cost for this case is the cost of HE plus the piping cost
4. Maintenance cost is assumed at 5% in of total investment cost

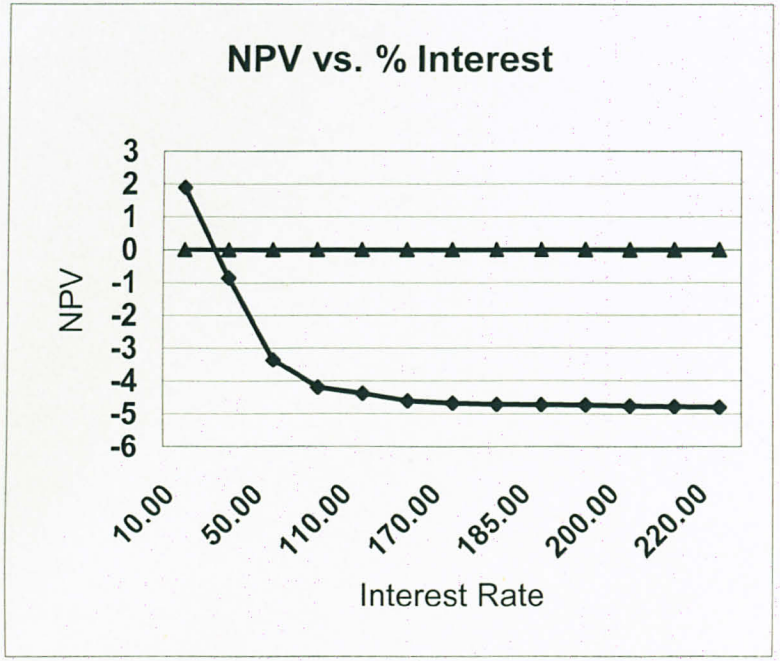
IRR is at 35% where NPV is zero.

NET PRESENT VALUE (NPV) AND INTERNAL RATE OF RETURN (IRR)

Description of the Project :LTSU

Year	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Savings		1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20	1.20
Maintenance		-0.26	-0.26	-0.26	-0.26	-0.26	-0.26	-0.26	-0.26	-0.26	-0.26	-0.26	-0.26	-0.26	-0.26	-0.26
Others																
Investment Cost	5.23															
Net Cash Flow		0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94	0.94

					NPV for 15 Years	
PV @ interest %	10.00	7.12	Million RM		NPV	1.89
PV @ interest %	20.00	4.38	Million RM		NPV	-0.85
PV @ interest %	50.00	1.87	Million RM		NPV	-3.36
PV @ interest %	90.00	1.04	Million RM		NPV	-4.19
PV @ interest %	110.00	0.85	Million RM		NPV	-4.38
PV @ interest %	150.00	0.62	Million RM		NPV	-4.61
PV @ interest %	170.00	0.55	Million RM		NPV	-4.68
PV @ interest %	180.00	0.52	Million RM		NPV	-4.71
PV @ interest %	185.00	0.51	Million RM		NPV	-4.72
PV @ interest %	190.00	0.49	Million RM		NPV	-4.74
PV @ interest %	200.00	0.47	Million RM		NPV	-4.76
PV @ interest %	210.00	0.45	Million RM		NPV	-4.78
PV @ interest %	220.00	0.43	Million RM		NPV	-4.80



Notes:

1. Figures are in Million RM
2. Assumed planned life of invesment is 15 years
3. Investment Cost for this case is the cost of HE plus the piping cost
4. Maintenance cost is assumed at 5% in of total investment cost

IRR is at 19% where NPV is zero.